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TECHNICAL AND ECONOMIC COMPARISON OF URANIUM RECOVERY FROM BIOLEACH SOLUTIONS BY THE USE OF BIOMASS, ION EXCHANGE AND REVERSE OSMOSIS

MCEL PROJECT No. 108

FINAL REPORT

MARCH, 1986

prepared for

DEPARTMENT OF ENERGY, MINES AND RESOURCES
CANADA CENTRE FOR MINERAL AND ENERGY & TECHNOLOGY

by

MELIS CONSULTING ENGINEERS LTD.
in association with
BEAK CONSULTANTS LIMITED
and
KILBORN (Saskatchewan) LTD.

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April 30, 1986

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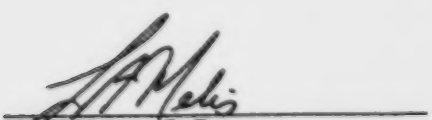
Dear Dr. Sanmugasunderam:

Melis Consulting Engineers Ltd. are pleased to submit ten (10) copies of the Final Report for the study entitled "Technical and Economic Comparison of Uranium Recovery from Bioleach Solutions by the Use of Biomass, Ion Exchange and Reverse Osmosis" which was completed for the Department of Energy, Mines and Resources under contract to the Department of Supply and Services.

We trust the information included in this report meets all your requirements and we remain available to answer any questions or discuss any aspects of the report at your convenience.

Melis Consulting Engineers Ltd., along with its sub-contractors, Beak Consultants Limited and Kilborn (Saskatchewan) Ltd., wish to express their sincere appreciation for the opportunity to be of service to the Biotechnology Section of CANMET in this component of their biotechnology studies and we look forward to providing further assistance in the future.

Respectfully submitted,
MELIS CONSULTING ENGINEERS LTD.



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TECHNICAL AND ECONOMIC COMPARISON
OF
URANIUM RECOVERY FROM BIOLEACH SOLUTIONS
BY
THE USE OF BIOMASS, ION EXCHANGE AND REVERSE OSMOSIS

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The assistance of Denison Mines Limited during the course of the study, in particular Mr. Andrew Marchbank, is greatly appreciated. Input from the Scientific Authority, Dr. V. Sanmugasunderam, throughout the study and in the review of the final report is also appreciated and hereby acknowledged.

TECHNICAL AND ECONOMIC COMPARISON
OF
URANIUM RECOVERY FROM BIOLEACH SOLUTIONS
BY
THE USE OF BIOMASS, ION EXCHANGE AND REVERSE OSMOSIS

1.0 SUMMARY

The Biotechnology Section of CANMET, Department of Energy, Mines and Resources, commissioned Melis Consulting Engineers Ltd. to evaluate the technical and economic feasibility of upgrading dilute bioleach solutions from underground leaching of uranium ores by the use of three alternative processes comprising ion exchange, biomass adsorption and reverse osmosis. This study was accomplished in association with Beak Consultants Limited, who were responsible for the literature review component of the study, and Kilborn Ltd., who assisted with the capital cost estimate of the three options evaluated.

Conceptual flowsheets and equipment layouts were detailed from process design criteria for each option which in turn were established from technical details obtained from the literature review as well as file information available from the study team. Using these flowsheets and layouts as a basis capital and operating cost estimates were prepared to afford an order-of-magnitude cost estimate for an ion exchange, biomass adsorption and reverse osmosis circuit for uranium recovery from underground bioleach solutions. An economic assessment of the three options identified for the study was then accomplished from these cost estimates.

The technical and economic results of the study are summarized in Table 1-1 for the three options. For the same solution throughput capacity the net annual uranium recovery is highest for the Reverse Osmosis Option but this higher capacity is reflected in a much lower concentration ratio which in turn signifies a lower potential for reducing

TABLE 1-1

URANIUM RECOVERY FROM BIOLEACH SOLUTIONS
TECHNICAL AND ECONOMIC COMPARISON SUMMARY
(FIRST QUARTER 1986 CANADIAN DOLLARS)

TECHNICAL	ION EXCHANGE	BIOMASS ADSORPTION	REVERSE OSMOSIS
Net % Uranium Recovery	95	82	99
Concentration Ratio	32:1	4.7:1	3.3:1
% Minewater Pumping Reduction	84	86	70
ECONOMIC			
Capital Costs			
Total $	1,674,400	3,736,600	3,680,500
$/annual m^3 processed	2.85	6.35	6.26
$/annual kg U_3O_8 recovered	19.98	51.83	42.15
$/annual lb U_3O_8 recovered	9.06	23.51	19.12
Operating Costs			
Annual $	374,300	3,519,820	539,640
$/$m^3$ processed	0.64	6.00	0.92
$/kg U_3O_8 recovered	4.46	48.83	6.18
$/lb U_3O_8 recovered	2.03	22.13	2.80
Total Costs*			
Total $	5,417,400	38,934,800	9,076,900
$/$m^3$ processed	0.92	6.62	1.54
$/kg U_3O_8 recovered	6.47	54.00	10.39
$/lb U_3O_8 recovered	2.93	24.49	4.71
Relative Total Cost Ratio			
based on m^3 processed	1.0	7.2	1.7
based on kg U_3O_8 recovered	1.0	8.3	1.6

* based on an assumed 10-year operating life

the volume of minewater to be pumped to surface in an underground leaching program. The highest concentration ratio, in terms of concentrated eluate production, would be obtained from an ion exchange circuit but the potential reduction in minewater pumping is limited by the requirement for a barren bleed in the ion exchange circuit needed for controlling sulphate levels in the recycled barren bioleach solution.

From a capital cost point-of-view the Ion Exchange Option is by far the most attractive option being less than half the cost of the Biomass Adsorption Option and the Reverse Osmosis Option.

The operating cost estimates have shown that the Biomass Adsorption Option would be approximately ten times greater in operating costs than the Ion Exchange Option while the Reverse Osmosis Option would be approximately one and a half times more costly to operate.

An order-of-magnitude comparison of total costs for each option was then made by calculating the total costs of each option for an assumed 10-year operating life at the rated capacity (70 m^3/h) used in the study and based on the respective uranium recovery for each option. On a total cost basis the Biomass Adsorption Option would still be significantly higher at seven times the cost of the Ion Exchange Option in terms of bioleach solution processed and eight times more costly in terms of uranium recovered. The total cost estimated for the Reverse Osmosis Option is approximately one and a half times more expensive than the Ion Exchange Option both in terms of solution processed and uranium recovered.

Based on current technology the technical and economic evaluation detailed in this report has shown ion exchange to be the most favourable processing option for in-situ uranium recovery from underground bioleach solutions. A comparison of ion exchange to the currently used method of lime precipitation for uranium recovery from minewater should

however be accomplished to assess the potential application of the lime precipitation alternative in an underground environment. This comparison would provide a true measure of the benefits or drawbacks of ion exchange in this application.

The Biomass Adsorption Option is currently unattractive due to the extent of process equipment required for a biomass adsorption circuit derived from available current technology for this option. Further developments in immobilization of biomass for producing a pelletized form would lead to improvements in its physical integrity thereby enabling the use of more simplified process equipment. The relatively low uranium loading efficiency of biomass along with its limited recycling potential and inherent high production cost also contribute significantly to the higher cost of this option. Consequently further research in these areas would be required to improve the viability of using biomass adsorption for uranium recovery from underground bioleach solutions.

The combination of inherent high costs of reverse osmosis units and solubility concerns in upgraded solutions leading to limited concentration capabilities make the Reverse Osmosis Option the least attractive of the three options in terms of future application for uranium recovery from bioleach solutions. It is expected that the high cost of reverse osmosis units would remain the same even if technical improvements could be realized for reverse osmosis.

COMPARAISON TECHNIQUE ET ECONOMIQUE DE LA RECUPERATION DE L'URANIUM A PARTIR DE SOLUTIONS DE LESSIVAGE BIOLOGIQUE PAR L'UTILISATION DE LA BIOMASSE, DE L'ECHANGE D'IONS ET DE L'OSMOSE RENVERSEE

1.0 RESUME

La section de biotechnologie de CANMET du Département d'Energie, Mines et Ressources Canada, chargea Melis Consulting Engineers Ltd. d'évaluer d'un point de vue technique et économique la possibilité de valoriser des solutions de lessivage biologique souterrain de minerai d'uranium en utilisant trois procédés possibles soit par l'échange d'ions, par l'adsorption de la biomasse et par l'osmose renversée. Cette étude fut réalisée en collaboration avec Beak Consultants Limited, responsables du volet de la révision de la documentation concernant l'étude, et Kilborn Ltd., qui ont aidé à établir l'estimation des coûts de capital des trois options évaluées.

Des shémas d'élaboration de fabrication et de montage de l'équipement furent établis à partir de critères basés sur l'étude de procédés pour chaque option qui, pour sa part, fut obtenue en fouillant la documentation de revues ou celle des dossiers fournis par les membres de l'équipe qui fit l'étude. Ces shémas de fabrication et de montage utilisés comme base, l'estimation des coûts de capital et de frais de fonctionnement furent préparés pour permettre une estimation de coût par ordre de grandeur pour un circuit d'échange d'ions, d'adsorption de la biomasse et d'osmose renversée pour la récupération d'uranium à partir de solutions de lessivage souterrain. Une évaluation du point de vue économique des trois options retenues pour l'étude fut alors faite à partir de l'estimation des coûts.

Les résultats techniques et économiques sont résumés au Tableau 1-1 pour chacune des trois options étudiées. Avec un débit de sortie de solution

TABLEAU 1-1

RECUPERATION DE L'URANIUM A PARTIR DE SOLUTIONS DE LESSIVAGE BIOLOGIQUE
RESUME DES COMPARAISONS TECHNIQUES ET ECONOMIQUES
(PREMIER TRIMESTRE 1986 DOLLARS CANADIENS)

COMPARAISON TECHNIQUE	ECHANGE D'IONS	ADSORPTION DE LA BIOMASSE	OSMOSE RENVERSEE
% net de la récupération de l'uranium	95	82	99
Rapport de la concentration	32:1	4,7:1	3,3:1
Réduction en % du pompage de l'eau	84	86	70
COMPARAISON ECONOMIQUE			
Coûts du capital			
Total en $	1.674.400	3.736.600	3.680.500
$/$m^3$ annuel traité	2,85	6,35	6,26
$/kg U_3O_8 annuel récupéré	19,98	51,83	42,15
$/lb U_3O_8 annuelle récupérée	9,06	23,51	19,12
Coûts d'opération			
$/an	374.300	3.519.820	539.640
$/$m^3$ traité	0,64	6,00	0,92
$/kg U_3O_8 récupéré	4,46	48,83	6,18
$/lb U_3O_8 récupérée	2,03	22,13	2,80
Total des coûts*			
Total en $	5.417.400	38.934.800	9.076.900
$/$m^3$ traité	0,92	6,62	1,54
$/kg U_3O_8 récupéré	6,47	54,00	10,39
$/lb U_3O_8 récupérée	2,93	24,49	4,71
Taux de total des coûts relatifs			
d'après le m^3 traité	1,0	7,2	1,7
d'après le kg U_3O_8 récupéré	1,0	8,3	1,6

* d'après une durée d'opération fictive de 10 ans

identique, la récupération annuelle nette d'uranium est plus grande selon le procédé d'osmose renversée, mais ce rendement supérieur implique un dosage de concentration inférieur, ce qui en fait signifie un potentiel plus bas en ce qui a trait à la réduction du volume de l'eau souterraine qui doit être pompée à la surface lors d'un lessivage souterrain. Le dosage de concentration le plus grand, en termes de produit élué concentré, serait obtenu à partir d'un circuit d'échange d'ions, mais la réduction possible du pompage de l'eau est restreinte par la nécessité d'utiliser un trop-plein stérile dans le circuit d'échange d'ions, nécessaire pour le contrôle des niveaux de sulfates dans la solution recyclée stérile de lessivage biologique.

D'un point de vue de coût du capital, l'option d'échange d'ions est de loin la plus intéressante, car celui-ci représente moins de la moitié du coût respectif des deux autres options étudiées.

Les estimations de frais de fonctionnement ont démontré que l'option d'adsorption de la biomasse coûterait environ dix fois plus que l'option d'échange d'ions en frais d'opération, contre une fois et demie de plus pour l'option d'osmose renversée.

Une comparaison par ordre de grandeur du total des coûts de chaque option fut alors faite pour chacune, ayant une période fictive de dix ans d'opération et un rendement donné de 70 m3/h, utilisé dans l'étude et établi d'après la récupération respective de chaque option. Basé sur un coût total, l'option d'adsorption de la biomasse serait encore manifestement plus onéreuse étant sept fois le coût de l'option d'échange d'ions en termes de solution traitée de lessivage biologique et huit fois plus coûteuse en termes d'uranium récupéré. En ce qui a trait au total des coûts pour l'option d'osmose renversée, celui-ci est d'environ une fois et demie plus élevé que l'option d'échange d'ions, tant au point de vue de la solution traitée que de l'uranium récupéré.

Selon la technologie actuelle, l'évaluation technique et économique élaborée dans ce rapport démontre que l'option d'échange d'ions est le

meilleur procédé pour la récupération in-situ de l'uranium obtenu à partir de solutions de lessivage biologique souterrain. Une comparaison de l'échange d'ions avec la méthode de précipitation par traitement à la chaux, utilisée présentement pour la récupération de l'uranium de l'eau de la mine, devrait par contre être réalisée afin de déterminer le potentiel de recyclage de cette possibilité en milieu souterrain. Cette comparaison permettrait d'évaluer à sa juste valeur les avantages et inconvénients de ce procédé.

L'option d'adsorption de biomasse est présentement peu attrayante dû à l'importance de l'équipement de traitement nécessaire pour un circuit d'adsorption de la biomasse selon les techniques du jour. De nouveaux progrès dans l'immobilisation de la biomasse afin d'en arriver à une production sous forme de granules en améliorerait la qualité physique, ce qui permettrait d'utiliser un équipement de traitement plus simple. Le rendement de la biomasse relativement bas du chargement d'uranium ainsi que son potentiel limité de recyclage contribue aussi grandement à son coût élevé, de telle sorte que des recherches supplémentaires dans ce domaine seraient nécessaires afin d'améliorer la viabilité de l'utilisation de l'adsorption de la biomasse dans la récupération d'uranium à partir de solutions de lessivage souterrain.

L'option de l'osmose renversée est la moins intéressante des trois possibilités en tant qu'utilisation future de la récupération de l'uranium à partir de solutions de lessivage biologique. La cause en est la combinaison des coûts élevés inhérents aux unités d'osmose renversée et aux problèmes de solubilité dans des solutions valorisées menant à des capacités de concentration limitées. Il est permis de croire que le coût élevé des unités d'osmose renversée demeurerait analogue même si d'autres améliorations techniques étaient apportées à l'osmose renversée.

TECHNICAL AND ECONOMIC COMPARISON OF URANIUM RECOVERY FROM BIOLEACH SOLUTIONS BY THE USE OF BIOMASS, ION EXCHANGE AND REVERSE OSMOSIS

2.0 INTRODUCTION

2.1 GENERAL

The recovery of uranium from underground minewater has been practiced in the Elliot Lake area uranium mines for a number of years and is currently being actively expanded by promoting the bacterial leaching of marginal uranium ores in-place underground (1,24). At present uranium recovery has been realized by pumping minewater to the surface mill facilities and precipitating the contained uranium with lime. The resultant precipitate is processed further in the main mill stream to achieve uranium recovery as final yellowcake product.

This natural bacterial leaching of uranium underground by acidic minewater can lead to higher uranium recoveries by promoting in-situ bioleaching of low-grade ores. Associated with an underground bio-leaching program in a uranium mine however are large quantities of leachate solution containing relatively low concentrations of uranium which need to be pumped to surface for uranium recovery.

The Biotechnology Section of CANMET, Department of Energy, Mines and Resources, has commissioned Melis Consulting Engineers Ltd.(MCEL) to investigate the technical and economic feasibility of upgrading dilute bioleach solutions underground by the use of three alternative processes comprising ion exchange, biomass adsorption and reverse osmosis. It is envisaged that in-situ upgrading of solution concentrations

underground would lead to decreased pumping requirements which would in turn lead to potential lower uranium recovery costs of an underground uranium leaching program.

MCEL contracted Beak Consultants Limited (BEAK) to provide assistance with the literature review component of the study to arrive at general conceptual flowsheets for each option. In this regard BEAK was assisted by technical experts in the three process areas, ion exchange, biomass adsorption and reverse osmosis, to minimize the literature search time as well as to provide input based on the most recent developments for each option. Kilborn Ltd. was also retained to assist with the capital cost estimating component of the study as part of an economic comparison of the three uranium recovery options.

This report details the findings of the study and provides a technical and economic comparison of the three process options to evaluate the viability of each option for upgrading uranium bioleach solutions.

2.2 SCOPE OF WORK

The defined scope of work for the study was to accomplish a literature search of the three process alternatives, encompassing in the large part recent research activities in each area, in order to derive suitable conceptual flowsheets for each option. From the detailed flowsheets a technical and economic comparison of the three alternatives was to be carried out along with an outline of experimental work that would be required to confirm any assumptions made in establishing the design criteria and to improve the technical and economic viability of the three options. The original Statement of Work is detailed in Appendix A.

The basic requirements established for the study were to accomplish the following for each option:

- literature review
- preparation of a design basis
- preparation of a conceptual flowsheet
- preparation of a conceptual equipment layout
- preparation of an order-of-magnitude capital cost estimate
- preparation of an operating cost estimate
- outlining experimental work for improving the possible viability of each option and for confirming any design assumptions.

Based on the information generated from the study a technical and economic comparison was to be completed to ascertain the viability of each option for upgrading underground uranium bioleach solutions. The findings of the study were to be detailed in a comprehensive report as well as presented at a seminar at the discretion of the Scientific Authority.

2.3 METHODOLOGY

As prime contractor for this study, Melis Consulting Engineers Ltd. engaged two sub-contractors to assist in the performance of the work. Beak Consultants Limited were engaged at the onset on the program to prepare a literature review of current technology related to ion exchange, biomass adsorption and reverse osmosis as applied to uranium recovery from bioleach solutions. In this regard they were assisted by Dr. V.I. Lakshmanan of Ontario Research Foundation (ion exchange), Dr. J.M. Sharer of the University of Waterloo and Senes Consultants Limited (biomass adsorption) and Dr. J. Dickson of McMaster University (reverse osmosis). Kilborn Ltd. were engaged later in the study to provide assistance with the capital cost estimate component of the study.

At the start of the program the study team met with representatives from the Biotechnology Section of CANMET, including the Scientific Authority, to initiate the study. At this meeting the basic assumptions of the study were agreed to in terms of facility location, study battery limits, process capacity, feedstock characteristics and the confidence level of cost estimates.

Following completion of the literature review a preliminary design criterion was established for each process option from which conceptual flowsheets were derived. Further discussions among the study team and with the Scientific Authority during preparation of the flowsheets finalized the most appropriate design basis which could be assumed for each option based on currently available technology.

From this conceptual design basis appropriate material balances were accomplished for each option and the most suitable equipment required for each process was selected from an underground installation point-of-view. The flowsheets were then completed and submitted to the Scientific Authority as part of the Stage I report in late December 1985. This report also included a summary of the literature review and

preliminary suggestions for experimental work which could be undertaken to substantiate any assumptions made in the study or to improve the viability of each process for recovering uranium from underground bioleach solutions.

Conceptual equipment layouts for each option were then prepared to provide quantity take-offs for detailing capital cost estimates. Concurrently budget-type prices were obtained from suppliers to afford the necessary information for completing capital cost estimates. Operating cost estimates were then detailed based on the conceptual design basis and flowsheet for each option. With all technical activities completed, efforts then concentrated on the preparation of the final report for submission at the end of March 1986.

2.4 BASIS OF STUDY

The basic assumptions of the study were agreed to among the study team and the Scientific Authority at the onset of the program. These formed the basis for detailing the three process options identified for potential uranium recovery from underground bioleach solutions. In the course of the study minor revisions were included to arrive at the most suitable conceptual design for ion exchange, biomass adsorption and reverse osmosis.

The main design assumptions are detailed below and the specific design parameters for each option are provided in the detailed description of each option in subsequent sections of the report. All units in this report are in the metric system. Consumption figures expressed as kg/day refer to kilogram per operating day based on a 350 day per year operation. Figures expressed as "kg/kg $U_3 O_8$" refer to kilogram per kilogram of recovered uranium applicable to each option.

Facility Location

Each process alternative was to be based on an existing uranium operation typical of uranium operations in the Elliot Lake region of northern Ontario. The conceptual processes for each option would be located in-situ in an underground environment. In the case of ion exchange, allowances would be made for elution of loaded resin and regeneration of fouled resin in a surface facility adjoining an existing mill. It is assumed that a decline from surface to the underground workings forms part of the mine facility at the uranium operation.

Study Battery Limits

The battery limits for the study were identified as bioleach solution as feedstock and the uranium-enriched solution generated by each

process as final product. For the Biomass Adsorption Option and the Reverse Osmosis Option the concentrated solution would report to the main mine sump for transfer to a surface mill facility and eventual uranium recovery in an existing mill circuit. In the Ion Exchange Option concentrated eluate from the surface elution circuit would report to the main eluate tank in the existing mill circuit. Any pre-treatment or auxiliary treatment required for successful operation of the proposed processes would be included as a component of the total process. Barren solution from each recovery circuit would be recycled to the underground bio-leaching area.

Process Capacity

The bioleach solution is assumed to originate from underground leaching of three blocks consisting of two mine stopes each. The average flowrate to the uranium recovery process would be 70 m^3 /h with an upside figure of 100 m^3/h as the maximum anticipated flow. A total of 350 operating days per year, 24 hours per day, has been assumed for the underground uranium recovery operation which equates to a plant availability of 96% relative to the total number of available days per year.

Feedstock Characteristics

The chemical and physical characteristics of the bioleachate feedstock assumed for the study are given in Table 2-1. The characteristics given represent estimated bioleach parameters in a once-through system (2). The uranium feed concentration has been assumed as 0.15 g/L U_3O_8. Iron and sulphate levels are assumed as 1 g/L Fe and 2.5 g/L SO_4 with peak values being 8 g/L and 15 g/L respectively.

Cost Estimates

The capital and operating cost estimates were to be carried out on an order-of-magnitude basis which would be consistent with the level of effort in a standard pre-feasibility study. All costs quoted in this study are in first quarter 1986 Canadian dollars.

TABLE 2-1

BIOLEACHATE SOLUTION CHARACTERISTICS

Parameter	Units	Median Concentration
U_3O_8	mg/L	150
ThO_2	mg/L	56
Y_2O_3	mg/L	20
Fe	mg/L	1,000 (8,000 peak)
Ti	mg/L	0.04
Cu	mg/L	1.0
Mn	mg/L	2.3
Zn	mg/L	3.6
Co	mg/L	1.6
Pb	mg/L	0.5
Ni	mg/L	1.2
Na	mg/L	30
K	mg/L	17
Ca	mg/L	186
Mg	mg/L	12
Ba	mg/L	0.1
Al	mg/L	43
PO_4	mg/L	25
Cl	mg/L	69
SO_4	mg/L	2,500 (15,000 peak)
NO_3-N	mg/L	180
NO_2-N	mg/L	79
NH_4-N	mg/L	45
SiO_2	mg/L	50-120
TSS	mg/L	10 (100 peak)
TDS	mg/L	2,700-3,500
pH		2.0-2.2
Temp - min.	^oC	8
- max.	^oC	17
Ra-226 (total)	Bq/L	7.14
Ra-226 (diss.)	Bq/L	5.14

3.0 ION EXCHANGE OPTION

3.1 LITERATURE REVIEW

3.1.1 Background

Ion exchange is a reversible process by which charged ions or molecules are exchanged between solid and liquid phases. The process can be modelled based on equations (isotherms) similar to equations developed by Langmuir and Freundlich for adsorption. In the case of uranium recovery ion exchange resins have been used for many years both for acidic and alkaline leach liquors(3).

The selection and type of ion exchange process to be used in a particular system is influenced by a number of factor, including:

- type of leaching (i.e. sulphuric acid, carbonate, etc.)
- ore characteristics
- concentration of uranium in the feed liquor
- occurrence of other ions in the feed liquor
- elution system and type of eluant to be used.

The dissolution of uranium by sulphuric acid or by sulphate bioleach solution is based on the following chemical equilibria(3):

$$UO_2^{2+} + SO_4^{2-} \longrightarrow UO_2SO_4$$
$$UO_2SO_4 + SO_4^{2-} \longrightarrow [UO_2(SO_4)_2]^{2-}$$
$$[UO_2(SO_4)_2]^{2-} + SO_4^{2-} \longrightarrow [UO_2(SO_4)_3]^{4-}$$

Adsorption of the uranyl sulphate complex on resin is based on the following reaction.

$$4R^+X^- + [UO_2(SO_4)_3]^{4-} \longrightarrow (R^+)_4\ UO_2(SO_4)_3 + 4X^-$$
(resin) (resin)

$(X^- = NO_3^-, Cl^-, HSO_4^-, SO_4^{2-})$

Elution of the loaded resin with a nitrate, chloride or sulphate solution occurs according to the following main exchange reaction and two side reactions:

$$(R^+)\ UO_2(SO_4)_3 + 4X^- \longrightarrow 4R^+X^- + UO_2^{2+} + 3SO_4^{2-}$$
$$\text{(resin)} \qquad\qquad\qquad\qquad \text{(resin)}$$

$$R^+HSO_4^- \quad X^- \longrightarrow R^+X^- + HSO_4^-$$

$$(R^+)_2SO_4 + 2X^- \longrightarrow 2R^+X^- + SO_4^{2-}$$

The mode of leaching, nature and concentration of chemicals used, and the nature of the ore all have an important bearing on the efficiency of resin adsorption.The effect of other ions on uranium recovery by ion exchange has been well described in Merritt's book on uranium metallurgy(3). Substances such as alkalis, alkaline cations, alumina and transition elements which do not form strong sulphate complexes do not have significant effects on uranium adsorption. The transition element ions which do not significantly effect uranium recovery are V^{2+}, Mn^{2+}, Fe^{2+}, Co^{2+}, Ni^{2+}, Cr^{2+} and Zn^{2+}. Anions such as HSO_4^- and SO_4^{2-} have a significant effect due to their influence in uranyl sulphate complex formation. At higher uranium and total sulphate concentrations lowering the acidity will decrease the HSO_4^- concentration, increase the SO_4^{2-} concentration, and increase the $[UO_2(SO_4)_3]^{2-}$ and $[UO_2(SO_4)_3]^{4-}$ concentrations. This will lead to an increase in uranium loading on the resin. However, with an increase in pH the iron sulphate complex loading can also increase.

An increase in uranium concentration can lead to an increase in uranium loading. The formation of uranyl sulphate complexes is promoted by increasing the sulphate concentration. However, bisulphate concentrations may increase under typical acid leach conditions causing a reduction in uranium loading at high sulphate concentrations.

Iron as Fe^{2+} does not have a significant effect on uranium loading. However, during acid leaching Fe^{3+} sulphate complexes are formed and occupy some of the active sites in the resin. These are displaced by the uranyl sulphate complex during the elution cycle. the effect of Fe^{3+} is generally small at a concentration less than 5 mg/L. At a pH

of greater than 1.8, however, ferric ions can decrease the uranium capacity of the resin.

Ions such as phosphate, arsenate and zirconium generally form insoluble salts which can poison the resin. Other resin poisons can include silica, molybdenum, polythionates, sulfur and titanium. Resin blinding can also occur due to the precipitation of calcium carbonate. Periodic regeneration of resin may consequently be required especially when processing minewater(4).

Common anions such as NO_3^-, Cl^-, or SO_4^{2-} can be used for elution. Sulphuric acid elution is less expensive and environmentally acceptable, but large amounts of alkali are required for neutralization.

3.1.3 Resin Type

Generally, the resin type used in uranium recovery is a strong base anionic exchange resin with a quaternary ammonium functional group. The wide use of this resin is due to the fact that it is highly effective in recovering anionic uranium complexes from acid sulphate solutions formed during sulphuric acid leaching.

Strong base resins have the advantages of high loadings and rapid elution kinetics as well as good physical characteristics(55). However, this type of resin loads other metal complexes, including iron, in addition to uranium(55). For this reason, the quality of the product stream is often poor and a weak base solvent extraction stage may be required after ion exchange in applications where high strength solutions are required for further processing.

The use of weak base resins has been proposed to provide an ion exchange process with the selectivity and hence product purity similar to that achieved by weak base liquid extractants(55). Two mills in the United States have used macroporous weak base ion exchange resins in fixed beds with brine as the eluant(55). The product quality, due to high resin selectivity for uranium in these systems, has allowed for

the direct precipitation of yellowcake from eluates. However, the loading capacity and kinetics are poorer than for strong base resins and thus longer cycle times are required. The poor loading kinetics therefore preclude the possibility of using weak base resins in systems treating dilute uranium solutions. In addition, silica fouling is a common problem for this type of resin requiring periodic caustic washing prior to reuse in the adsorption stage.

Resin density is an important parameter when considering the employment of an upflow continuous ion exchange system. The solids content of the slurry and velocity of the flow are dependent on the density or size of the resin particles. During operation of these systems less dense or smaller resins would be washed out of the column resulting in high resin losses.

3.1.4 Mode of Operation

There are several ion exchange systems which can be used as the method of operation for ion exchange recovery of uranium.

Static Beds Systems

Original ion exchange plants used the static bed system of operation with downflow of clarified liquor through a series of packed beds. A static system can be of fixed bed type, where adsorption and elution occurs in the same column or it can be of the moving bed type where resin is transferred to separate columns for adsorption and elution. In the Elliot Lake area the Quirke operation utilizes a fixed bed system while the Panel and Denison operation utilizes a moving bed system.

Static bed systems have particular application in treating low-grade solutions where the adsorption cycle is long compared to elution(12).

Continuous Counter-Current Systems

In upflow continuous counter-current systems (CIX) the resin and feed solution move in opposite directions in the adsorption column with the

loaded resin being moved to a separate column for elution. The major advantage of this system is the ability to handle unclarified liquors or dilute pulp. Other advantages include lower resin inventory than static beds, more compact design and reduction in reagent usage(55).

There are a number of continuous ion exchange systems which have been developed. These include the Chem-Seps System (5), the Himsley Column (6,7,8,9,10), the Darcy George Column (11), the NIMCIX System(12,13), and the Deep-Fluidized Bed System(14,15,16).

The Chem-Seps System has been used at Kennecott Corporation's Bingham Canyon Mine to recover uranium from copper mine run-off but its greatest application is in municipal water softening. The Himsley multi-stage fluidized bed ion exchange system has been employed for uranium recovery in Canada at the Agnew Lake Mine near Espanola, Ontario and is currently used at Rio Algom Ltd.'s Stanleigh Mine in Elliot Lake, Ontario. Only pilot tests have been reported for the Darcy George Column(11). The NIMCIX Column has been found to work satisfactorily in pilot scale tests at Hartebeestfontein, South Africa, in full-scale operation at Blyvooruitzicht, South Africa, and at United Nuclear-Homestake Partner's Mill in Grants, New Mexico. A deep fluidized bed column was tested at Eldorado's Beaverlodge Mine to recover uranium for minewater(14).

Although continuous ion exchange systems have found various applications for uranium recovery, past discussions with various operators have indicated that these systems are not trouble free and that control requirements may be too excessive for an underground application. Their use of upflow fluidized bed or expanded beds is an attractive consideration however since this eliminates the need for clarification of feed solution as required for downflow static beds.

Other Systems

Upflow counter-current tank systems can be used for uranium adsorption where unclarified solution is pumped upflow through a series of tanks.

Resin is advanced from tank to tank and loaded resin is eluted in standard downflow counter-current columns. The major advantage of this system is the large cross-sectional area available which permits the processing of large volumes of solutions.

Resin-in-Pulp systems have been used in uranium leach plants where dissolved uranium is recovered by contacting resin in pulp slurries(3).

The Recoflow ion exchange system(17) designed by Eco-Tec Ltd. is characterized by the employment of fine-particle size resins (100-200 mesh), counter-current elution, short column heights (7 to 60 cm), fast flow and short cycles. It could prove to be an attractive alternative for recovering uranium from dilute minewater solutions since the fine resin size leads to faster reaction kinetics and thus higher flowrates.

Summary

Considering the in-situ conditions of an underground facility, where any uranium recovery process must be forgiving, inexpensive and simple to operate, a simple three-column cyclic operation using a static bed system would be the most suitable method of operation. Adsorption of uranium in partially fluidized beds under upflow conditions would provide a simple means of operation while maintaining the flexibility of processing feed solutions with variable suspended solids content.

3.2 GENERAL PROCESS DESCRIPTION

The conceptual flowsheet and material balance for ion exchange uranium recovery from bioleach solutions is provided in Drawing No. 100-10-1. A conceptual layout of the required equipment is shown in Drawing No. 100-20-1. The design criteria used in establishing this flowsheet and layout are itemized in Section 3.3. The envisaged process would lead to a 95% uranium recovery of the uranium contained in the bioleach solution and result in a solution concentration ratio of 32 to 1 based on the uranium concentration in the concentrated eluate liquor relative to the uranium concentration in the bioleach solution feed.

The process includes two parallel sets of ion exchange columns with two columns in each set being in an adsorption mode at any one time and the resin in the third column either being eluted or backwashed prior to elution. Upflow operating conditions have been assumed for the adsorption columns in order to accomodate any suspended solids in the bioleach feed solution. Downflow could however be utilized if the clarity of bioleach solution feed proved to be acceptable. Loaded resin would be transported to the surface mill via a decline for elution with nitrate solution. A resin regeneration facility adjoining the surface mill is also included for treating poisoned resin.

Bioleach solution in the underground mining stopes would be collected in an ion exchange feed tank and pumped to two parallel sets of ion exchange columns. The flow would be split evenly between the two sets and fed in series through two columns for recovery of uranium. The specific flowrate through the columns would be sufficient to maintain a 20% bed expansion under upflow conditions.

Once the resin in the lead column was loaded to approximately 48 kg/m^3 U_3O_8, which has been calculated to take two and a half days based on 119.7 kg/day uranium feed to each column and a total of 6.2 m^3 of resin per column, it would be backwashed with barren solution to remove any entrained solids. The second column would then become the lead column for adsorption and the third column, containing eluted resin, would be placed in service as the second adsorption column in series.

Barren solution exiting from both sets of ion exchange columns would be collected in a tank and recycled to the underground bio-leaching area. A barren bleed stream would be discharged directly to general mine drainage since recycling the total flow of barren solution would lead to continuous increase in sulphate concentration of the leachate to levels greater than the 15 g/L SO_4 established as the upper limit for the study. The net minewater make-up to the barren recycle tank would thus be 11 m^3/h.

In order to avoid the need to transfer large quantities of sulphuric acid underground it has been assumed that elution of loaded resin would be carried out in a separate elution column located on surface near an existing mill facility. Nitrate elution has been assumed for the existing mill circuit and hence for the bioleach loaded resin. Loaded and eluted resin would be transported by a mobile tanker via an existing decline from surface to the underground leach area. The concentrated eluate would be discharged to the mill eluate storage tank at an average concentration of 4.8 g/L U_3O_8. Higher eluate concentrations (10-12 g/L U_3O_8) could be achieved if the resin was eluted with recycle of eluate solutions as practiced in the existing mill circuit but a value of 4.8 g/L U_3O_8 has been used in the study since the bioleach ion exchange circuit was considered as a stand-alone circuit for comparison purposes.

Resin fouling by silica and other elements contained in the bioleach solution would lead to the requirement for a resin regeneration facility. Due to the large amounts of sulphuric acid required for this it has been assumed that resin regeneration would take place annually on surface. A single column would be required for this purpose with 30% sulphuric acid being used as regeneration solution. Spent acid would be pumped to the mill leaching circuit to make use of the available acidity in the spent acid. Resin would be transported to surface in a mobile tanker via the mine decline. Portable resin transfer pumps would effect the transfer of resin between the columns and the mobile tanker.

Reagent requirements for this process only include nitric acid for eluate make-up and sulphuric acid for resin regeneration. Fresh eluate for eluting loaded resin would be available from an existing mill supply. The quantity of nitric acid make-up to the existing mill eluate tank required for eluting the bioleach loaded resin is estimated as 3.5 kg/kg U_3O_8 based on operating data for a nitric acid elution system(23).

A separate tank would be provided for preparing 30% H_2SO_4 solutions. Concentrated sulphuric acid would be taken from an existing mill supply and diluted with process water in a static mixer. The annual sulphuric acid requirement for regeneration has been calculated as 154 tonnes (93% H_2SO_4) or an average of 1.84 kg/kg U_3O_8.

3.3 Design Criteria

The design criteria established for the ion exchange process flowsheet and associated conceptual layout were obtained generally from file data and operating data from current ion exchange operations. The criteria details and assumptions used for the Ion Exchange Option are detailed below.

3.3.1 General

Bioleach solution flow (m^3/h) - average	- 70
- maximum	- 100
Bioleach solution analysis (g/L U_3O_8)	- 0.15
Net ion exchange uranium recovery (%)	- 95
Concentration ratio	- 32:1
Uranium recovery - kg/day U_3O_8	- 239.4
- kg/year U_3O_8	- 83,790
Operating days per year	- 350
Operating system	three-column, cyclic operation

3.3.2 Adsorption

Cross-sectional area per column (m^2)	- 3.1
Number of columns per set	- 3
Number of sets	- 2
Column diameter (m)	- 2.0
Column height (m)	- 3.25
Materials of construction	- rubber-lined steel
Operating mode	- upflow,two-stage
Resin	- Dowex 21 k or equivalent
Adsorption efficiency (%)	- 95
Bed expansion (%)	- 20
Upflow rate (m/h)	- 11
Feed rate per set (m^3/h)	- 35
Specific flowrate ($m^3/h/m^2$)	- 11
Barren bleed (m^3/h)	- 11

Barren analysis g/L U_3O_8	- 0.0075
Resin static bed height (m)	- 2.0
Resin bed volume per column (m^3)	- 6.2
Resin loading (kg/m^3 U_3O_8)	- 48
Loading time (days)	- 2.5
Feed tank capacity (min)	- 5
Barren tank capacity (min)	- 5

3.3.3 Backwash

Operating mode	- upflow
Bed expansion (%)	- 50
Upflow rate (m/h)	- 17
Backwash flow per column (m^3/h)	- 53
Specific flowrate (m^3/h/m^2)	- 17
Backwash duration (min)	- 5
Resin trap retention time (min)	- 10

3.3.4 Elution

Operating mode	- downflow
Loaded resin volume per column (m^3)	- 12.4
Resin loading (kg/m^3 U_3O_8)	- 48
Uranium load per column (kg U_3O_8)	- 598.5
Elution frequency per column	- every 2.5 days
Number of bed volumes for elution	- 10
Eluant solution	- 62 g/L NO_3^-
Volume of eluant per column (m^3)	- 124.4
Flowrate (m^3/h)	- 20
Specific flowrate (m^3/h/m^2)	- 4.1
Elution cycle time (h)	- 6.2
Average eluate concentration (g/L U_3O_8)	- 4.8
Concentration ratio	- 32:1
Acid consumption (61.4% HNO_3)	- kg/kg U_3O_8 - 3.5
	- kg/day - 838
	- m^3/day - 0.6

Elution column diameter (m)	- 2.5
Elution column height (m)	- 3.75
Nitric acid supply	- from mill supply
Water rinse bed volume	- 1
Water rinse flowrate (m^3/h)	- 20
Water rinse discharge point	- mill eluate tank

3.3.5 Regeneration

Frequency	- annually
Fouled resin analysis (% Ash)	- 15-20
Regenerated resin analysis (% Ash)	- 1
Regenerating solution	- 30% H_2SO_4
Number of bed volumes	- 7
Acid regeneration time	- 8
Total flow per column (m^3)	- 87.1
Flowrate (downflow) (m^3/h)	- 11
Specific flowrate ($m^3/h/m^2$)	- 2.2
Regenerated resin volume per year	- 37.3
Total volume of 30% H_2SO_4 per year (m^3)	- 261.3
Regeneration acid consumption (93% H_2SO_4)	
- m^3/year	- 84.3
- t/year	- 154.3
- kg/kg U_3O_8	- 1.84
Spent acid discharge point	- mill leaching circuit
Water rinse flowrate-upflow (m^3/h)	- 83
Upflow rate (m/h)	- 17
Number of bed volumes	- 7
Water rinse time (h)	- 8
Water rinse discharge point	- mill tailings pumpbox
Regeneration column diameter (m)	- 2.5
Regeneration column height (m)	- 3.75

3.4 CAPITAL COST ESTIMATE

3.4.1 Summary

The cost summary presented in Table 3-1 provides the estimated capital costs of the Ion Exchange Option for recovering uranium from underground bioleach solution with a solution feed rate of 70 m^3/h. These estimated costs total $1,674,400 in first quarter 1986 Canadian Dollars and provide for all costs of the ion exchange facility including underground adsorption and surface elution and regeneration.

The estimate includes:

- underground excavation and concrete pad
- surface building complete with services
- underground adsorption circuit
- surface elution and regeneration circuit with reagent make-up
- mobile tanker for resin movement
- indirect costs of construction
- engineering, procurement, and construction management
- capital spares inventory
- initial circuit resin inventory
- contingency allowance.

The capital cost estimate details for the Ion Exchange Option are provided in Appendix B.

The total unit capital costs of the option can be interpreted as follows:

- $2.85 per annual cubic meter of bioleach solution processed
- $19.98 per annual kilogram of U_3O_8 recovered
- $9.06 per annual pound of U_3O_8 recovered.

These cost estimates do not include any provision for pre-production construction interest as the total value of this interest is dependent

TABLE 3-1

ION EXCHANGE OPTION
CAPITAL COST ESTIMATE

Direct Costs	$
Surface Building	90,292
Underground Excavation	108,264
Equipment	432,948
Mechanical	80,500
Piping	138,600
Electrical	199,175
Instrumentation	41,100
Total Direct Costs	1,090,879
Indirect Costs	
Construction Indirects (8%)	87,270
Engineering, Procurement, and Construction Management (12%)	130,905
Capital Spares	16,100
Resin Inventory	197,000
Sub-Total	1,522,154
Contingency (10%)	152,215
TOTAL ESTIMATED CAPITAL COSTS	1,674,369
	say, 1,674,400

on the method of financing and the equity or credit position of the specific investor. These pre-production interest costs can have a significant impact on the total costs of any project and this should be kept in mind when reviewing the cost estimates given in this report.

3.4.2 Basis of Estimate

The capital cost estimate is based on the following information:

- flowsheet and conceptual layout prepared by Melis consulting Engineers Ltd.
- budget quotations on major equipment
- pricing data from Melis Consulting Engineers Ltd. and Kilborn Ltd.'s files and records to arrive at factors for mechanical, piping, electrical and instrumentation costs. These factors have been adjusted to reflect the degree of complexity of the proposed process and the relative cost of unit equipment.

For purposes of this study it has been assumed that:

- all construction services would be tendered by fixed price contract
- the average labour rate is $35.00 per hour based on a 40-hour week. This labour rate includes provision for direct wages, payroll burdens and benefits, small tools and consumables, site preparation and the contractor's overhead and fee
- construction labour force is available in the immediate surrounding area
- the cumulative price factor applied to base prices for process equipment is eighteen percent,which includes:
  - provincial sales tax - 7%
  - freight allowance - 5%
  - miscellaneous minor materials allowance - 5%
- all costs are reported in first quarter 1986 Canadian dollars.

No allowances have been made in the capital estimate for:

- escalation beyond the first quarter 1986
- pre-production construction interest
- working capital
- production equipment or materials
- start-up assistance
- cost of pre-appropriation studies and development work
- owners' costs such as owners' representative
- general property facilities such as administration offices
- pumping bioleach solutions from the underground leaching areas to the treatment facility
- pumping of barren bleed to surface with the general minewater
- decommissioning.

3.4.3 Methodology

Equipment prices were obtained on an individual basis from budget quotations and historical data. Mechanical installation costs, including field material, were factored from the equipment costs. Piping, electrical and instrumentation costs were factored from the total installed mechanical costs. Building structure costs for each area were based on volumes and volume unit prices derived from historical data. The factors used in deriving the total installed mechanical costs have been adjusted to reflect the complexity and relative cost of unit equipment for the proposed process.

The allowance for construction indirects has been estimated as eight percent of the total direct costs. It includes:

- temporary services
- permits, insurance and bonds
- winter work (for the surface building)
- rental of mobile equipment
- on-site construction supervision facility
- pre-construction geotechnical investigations.

The engineering, procurement and construction management function has been estimated as twelve percent of the direct costs. As well as design engineering, this includes provision for project management, cost control, scheduling, contract administration and inspection services.

An allowance of five percent of home office costs has been included for capital spares inventory. In addition a resin inventory allowance has been included based on a resin price of $\$5,300/m^3$($\$150/ft^3$).

The contingency allowance provdes for additions and unidentified items within the stated scope of work but it is not intended to cover major changes in scope.

3.5 OPERATING COST ESTIMATE

3.5.1 Summary

The ion exchange process operating costs, based on the design basis detailed in Section 3.3, total $374,300 annually in first quarter 1986 Canadian dollars, including a five percent contingency allowance. This annual operating cost is equivalent to $\$0.64/m^3$ of bioleach solution processed or $\$4.46/kg\ U_3O_8$ ($\$2.03/lb\ U_3O_8$) based on a total of 83,790 kg U_3O_8 (184,722 lbs U_3O_8) of recovered uranium per year for the Ion Exchange Option.

A summary of the operating costs for a 350-day per year operation is presented in Table 3-2 with cost details given in Tables 3-3 and 3-4. The operating costs include all costs associated with the recovery of uranium from underground bioleach solutions by the proposed ion exchange process. The processing costs extend from receipt of the bioleach solution underground to the production of concentrated eluate on surface. These costs include manpower, reagents and supplies, power, heating for the surface building, and the cost of analysis for circuit operation. It does not include any costs associated with the actual underground leaching nor does it include any costs associated with downstream processing of the uranium in the surface milling circuit.

3.5.2 Manpower

The annual manpower costs detailed in Table 3-3 include the manpower specifically required for the operation of the adsorption circuit underground and for operation of the elution and regeneration circuits on surface. The average annual total manpower is estimated at four men for meeting operation and maintenance requirements. A twenty-five percent payroll burden has been assumed in addition to the indicated salary rates.

TABLE 3-2

ION EXCHANGE OPTION
ANNUAL OPERATING COSTS

Item	Annual $	$/m^3	$/kg U_3O_8	$/lb U_3O_8
Manpower	179,100	0.30	2.13	0.97
Reagents and Supplies	134,400	0.23	1.60	0.72
Power	19,800	0.035	0.24	0.11
Heating	3,200	0.01	0.04	0.02
Analysis	20,000	0.035	0.24	0.11
Sub-Total	356,500	0.61	4.25	1.93
Contingency (5%)	17,800	0.03	0.21	0.10
TOTAL	374,300	0.64	4.46	2.03

TABLE 3-3

ION EXCHANGE OPTION
ANNUAL MANPOWER COSTS

Employee	Requirements	Number	Rate	Annual $
Supervision-Operation	0.5 man/day	0.5	$3,500/month	21,000
-Maintenance	0.1 man/day	0.1	$3,500/month	4,200
Operator-Underground	0.5 man/shift	2.0	$16.00/hour	67,200
-Elution	250 mandays/year	0.8	$16.00/hour	32,000
-Regeneration	20 mandays/year	0.1	$16.00/hour	2,600
Maintenance	10 mandays/month	0.5	$17.00/hour	16,300
Sub-Total		4.0		143,300
Payroll Burden (25%)				35,800
TOTAL ESTIMATED MANPOWER COSTS				179,100

TABLE 3-4

ION EXCHANGE OPTION
ANNUAL REAGENTS AND SUPPLIES COSTS

Item	Unit Cost($/kg)	Consumption kg/kg U_3O_8(1)	Consumption kg/year	Annual $
Nitric Acid (61.4%)	0.114	3.5	293,265	33,460
Sulphuric Acid (93%)	0.0385	1.84	154,300	5,940
Resin Losses(2)				10,000
Maintenance Supplies				50,000
Operating Supplies				30,000
Laboratory Supplies				5,000
TOTAL				134,400

Notes:

(1) Reagent consumptions are based on 83,790 kg of recovered uranium (U_3O_8) per year

(2) Annual resin loss is estimated as 5% of the original inventory.

The supervision allowance for operations is based on a supervisor spending fifty percent of his time on dayshift monitoring the operation of the ion exchange circuit. Maintenance supervision would be less with a supervisor spending ten percent of his time in the area.

Operating manpower for the underground adsorption circuit is based on one man spending fifty percent of his time for this circuit, twenty-four hours per day, 350 days per year. A total of 250 mandays per year has been allocated for resin elution which includes the manpower for transporting resin between underground and the surface elution facility. A total of 20 mandays would meet the annual resin regeneration requirements. A maintenance allowance of 10 mandays per month has been assumed for maintenance of the ion exchange circuits.

3.5.3 Reagents and Supplies

The cost of reagents and supplies detailed in Table 3-4 is based on consumption rates given in the ion exchange design basis and the delivered price of reagents to the facility. The annual consumption is based on 83,790 kg of recovered uranium per year.

The nitric acid cost is the cost of acid for making up fresh eluate reuired for elution of the bioleach loaded resin. The indicated sulphuric acid cost of regeneration is included as an operating cost although discharge of the spend acid to the mill leaching circuit would lead to utilization of the acid for leaching.

A resin loss allowance is included due to the transfer and transportation of loaded and eluted resin between underground and surface. Lump sum estimates have been included for laboratory, maintenance and operating supplies sufficient to meet the requirement for operation of the ion exchange circuits.

3.5.4. Power

The total power demand for operating the ion exchange circuits is estimated at 100 kW. The total annual power cost is thus estimated at $19,800 assuming an 85% power efficiency factor and a power cost of $0.02/kWh for a 350 day per year operation.

3.5.5 Heating

The annual heating cost of $3,200 is the estimated cost for heating the surface building housing the elution and regeneration circuits.

3.5.6 Analysis

The lump sum estimate of $20,000 for analytical costs, which would be sufficient for approximately 2,000 analyses per year, would meet the requirements for daily metallurgical control of the adsorption and elution circuits.

3.6 TECHNICAL EVALUATION

3.6.1 Discussion

Ion exchange resins have been used for recovery of uranium for many years both from acidic and alkaline leach liquors. As such the process flow detailed in Drawing No. 100-10-1 is generally based on well established practice and industrial experience and can be considered as a practical approach to uranium recovery from bioleach solutions by ion exchange.

A net ion exchange uranium recovery of 95% has been shown to be a realistic estimate both from previous testwork on uranium recovery from minewater or dilute liquids (5,11,12,14,15,18,19) and from actual plant data (4,20). In fact recoveries greater than this have been achieved but a value of 95% was felt to be a reasonable estimate for the purpose of this study.

A total of 350 operating days per year, 24 hours per day, has been assumed which equates to a plant availability of 96%. Downtime would be expected to be minimal for this type of operation and at least one week per year would be used for resin regeneration.

A three-column cyclic operation is a standard installation for a static bed ion exchange system. Two stages of adsorption would ensure adequate recovery efficiencies with upflow being taken as the best mode of operation to avoid the need for bioleach solution clarification ahead of ion exchange. A static bed system under upflow conditions has been selected as the simplest means of operation for an underground facility while maintaining the flexibility of processing feed solutions with variable suspended solids content.

A typical resin type for uranium recovery would be Dowex 21 K or its equivalent. Bed expansion and specific flowrate values are based on the physical characteristics of the Dowex resin (21).

A barren bleed has been included in the design to control sulphate levels in the bioleach solution and to mitigate potential resin poisoning effects of other ions present in solution (18). An ion exchange barren analysis of 0.0075 g/L U_3O_8 is indicated (based on 95% adsorption efficiency) although values as low as 0.002 g/L U_3O_8 would be possible. It is noted however that high uranium recovery efficiencies are not critical to this process since the barren will be recycled to the underground leaching areas.

The resin bed volume for each column is based on a resin loading of 48 kg U_3O_8/m^3 which will allow a 2.5-day loading time at the flowrate and feed analysis assumed for the study. A resin loading of 48 kg/m^3 is equivalent to 3 lbs/ft^3 which was the loading obtained on resin during recovery of uranium from minewater at the Quirke Mine operation in 1975 and 1976 (4). Since it is based on actual operating data it is felt to be a realistic estimate of resin loading that can be anticipated for the ion exchange option of uranium recovery from bioleach solutions. In comparison typical uranium loadings from acid solutions in conventional uranium circuits range from 4 to 5 lbs/ft^3(3).

A backwash stage has been assumed as necessary to remove any entrained solids which may build-up over the 2.5-day loading time. A specific flowrate of 17 $m^3/h/m^2$ will provide a bed expansion of 50% (21) which is adequate for efficient backwashing. A resin trap will avoid the loss of resin during upset conditions. Barren recycle would be used for backwashing.

Elution of loaded resin on surface in a facility adjoining an existing uranium mill has been selected for the Ion Exchange Option to avoid the need of transferring substantial amounts of acid underground for elution of resin in-situ. Transporting of resin via a mobile tanker using an existing underground decline from surface would not be problematic since resin transfers take place regularly in moving bed systems. Nitrate elution has been selected to conform with the assumed nitrate elution circuit of the main mill process.

Downflow elution is standard practice in industry and is practiced in the Elliot Lake operations (4,22,23). The elution conditions given are based on operating practice (23). The eluate concentration of 4.8 g/L U_3O_8 is the average concentration over the 10 bed volumes of eluate. Standard methods of eluate recycle would lead to concentrations approaching 10-12 g/L U_3O_8 but an average value has been assumed for this study.

Resin fouling proved to be a problem in the Quirke operation on uranium recovery from minewater (4) and is generally a problem when processing liquid of varying chemistry (3). Regeneration of the resin was therefore necessary to reduce resin fouling from levels of 15 to 20% ash down to levels of approximately 1%. A total of 7 bed volumes of 30% H_2SO_4 solution was sufficient to achieve this regeneration efficiency (4). The same regeneration procedure used in the Quirke operation has been included in the process flowsheet for this study. A water rinse is included to remove strong acid from the resin prior to transferring the regenerated resin back to the ion exchange columns underground.

Since the regeneration step is a once-through process, and the total quantity of acid required for this is significant, it would be necessary to carry out resin regeneration on surface adjacent to the mill facility. This would enable the re-use of spent acid in the mill leaching circuit and thus minimize the direct costs for regeneration.

3.6.2 Recommendations

Although ion exchange technology is well established the following items have been identified as avenues of future experimental work when considering recovery of uranium from dilute bioleach solutions:

- a brief evaluation of available resins in terms of uranium adsorption and loading efficiency

- identification of the physical characteristics of the most suitable resin for design purposes
- the effect of gradual increases in the concentration of cations and anions caused by the recycle of barren liquor to the underground leaching areas
- the effect of Fe^{3+}, SO_4^{2-}, and HSO_4^- ions on uranium adsorption
- the effect of various elements present in bioleach solutions on resin poisoning
- determination of the operating life of resin between regeneration stages and uranium loading expectancies relative to increasing levels of resin fouling
- optimization of regeneration conditions for fouled resin
- confirmation of elution conditions and efficiencies and investigation of possible eluates which could be used in an underground environment
- the effect of minewater temperature changes on uranium adsorption kinetics
- the efficiencies of ion exchange with respect to the recovery of yttrium from underground bioleach solutions.

4.0 BIOMASS ADSORPTION OPTION

4.1 LITERATURE REVIEW

4.1.1 Background

Biomass adsorption makes use of the fact that certain microorganisms have the ability to selectively sorb various constituents from solution. Subsequent separation and elution of loaded or sorbed biomass offers a means of scavenging or concentrating elements of interest thereby leading to potential applications in hydrometallurgical processes. The potential application of biotechnology in the mineral industry has been well summarized by Campbell et al (24).

The capability of microorganisms to extract metal ions from aqueous solutions has been known for over thirty years(25). Most studies have ususally addressed this phenomenon from a perspective of environmental toxicity. For example, fresh water and marine algae are known to accumulate uranium(26,27,28,29) and other heavy metals to levels that may be toxic to higher aquatic species(30). Also the bioconcentration of toxic heavy metals in sewage sludge leads to restrictions for disposal on agricultural land. The natural adsorption of uranium from groundwater by peat has also been identified(31).

While concerns about environmental safety were the original impetus for research, several recent studies have dealt with the potential exploitation of microorganisms for bioconcentrating metals from hydrometallurgical process liquids(32). Considering that uranium leaching is a biologically mediated process involving Thiobacillus ferrooxidans and related species, it is a logical extension to examine biomass further for uranium recovery.

4.1.2 Details

Microorganisms use various organic macromolecules on their external surfaces (polypeptides, proteins, polysaccharides) to capture essential metal ions from dilute solutions. Unlike other essential metal ions, the uranyl ion and its derivatives are not known to function in any metal-enzyme complex. Yet, the bioadsorption of uranium has been demonstrated in a variety of microorganisms including bacteria, fungi, yeasts, and algae. This phenomenon has been observed in both natural systems and under artificial conditions in the laboratory.

The data base on uranium bioaccumulation in freshwater biota has been reviewed by Blaylock(33). The most complete set of data available is for Lake Michigan. In this lake, the U^{238} bioaccumulation factor (BF) ranges from 88-145 mg/g (fresh weight) in phytoplankton to 0.3-0.7 mg/g (fresh weight) in piscivorous and omnivorous fish. Differences in bioaccumulation exist among the members of the phytoplankton community. Sakaguchi et al(26,27) have observed that the uptake of uranium by fresh water microalgae from 1 mg/L uranium solution (pH 8.0) is the highest in Chlorella regularis cultures (3.93 mg/g dry weight) followed by various Chlamydomonas (2.33-3.40 mg/g dry weight) and Scenedesmus (0.08-1.92 mg/g) species.

A definitive comparison of other protists with algae is difficult and tentative at best, since conditions for bioadsorption varied among the investigators. In general, actinomyces (Streptomyces levoris, Streptomyces viridochromogenes) appear to have the highest uranium bioadsorption capacities(28,29) followed by bacteria such as Pseudomonas aeroginosa (28,29,34), Zooglea ramigera(35), municipal sludge(36), Thiobacillus ferrooxidans(37), the yeast Saccharomyces cerevisiae(28, 29) and the fungi including Aspergillus niger and Aspergillus terreus (36), in decreasing order.

Fungi belonging to genus Rhizopus and Penicillium have been found to be the most notable exceptions of the above order(36) in as much as the

maximum uranium adsorption capacities of P. chrysogenum (165 mg U/g biomass) and R. arrhizus (180 mg U/g biomass) exceeded all other species.

Senes Consultants Limited(38) conducted extensive laboratory studies on uranium recovery using biomass. Selected conclusions from this work were as follows:

- the fungus Rhizopus arrhizus is simple to cultivate in aerobic culture, it produces high yields and is easily harvested
- bioadsorption capacity is a strong function of pH. Optimum levels that maximize adsorption capacity and uranium concentration in process solutions would be in the 3.5 to 4 range
- the presence of contaminants such as iron, sulphate and heavy metals reduces bioadsorption capacity
- sodium bicarbonate is an effective stripping agent for the biomasses tested
- up to 10 adsorption/desorption cycles are feasible before biomass replacement
- simple isotherms such as the Freundlich and Langmuir isotherm described the experimental data.

A limited bench scale test using Rhizopus arrhizus with uranium bioleach solution was completed by Zenon Environmental Inc. for Denison Mines Limited(39) with results that were generally consistent with the SENES conclusions. Uranium uptakes of 48 mg/g and 5 mg/g of dry biomass were observed at test pH's of 4.0 and 2.8 respectively. This pH adjustment step is therefore necessary to increase solution pH to a range which allows reasonable loading.

The limited amount of development work which has been done with biomass provides only basic data for preparation of a conceptual flowsheet for uranium recovery from bioleach solutions. Further development activities will certainly lead to improvements but this study can only assume current technical data which has been gathered from the literature review.

4.2 GENERAL PROCESS DESCRIPTION

The detailed flowsheet and material balance for uranium recovery from bioleach solutions by biomass adsorption is provided in Drawing No. 100-10-2. A conceptual layout of the equipment required for this process is shown in Drawing No. 100-20-2. The design criteria used in establishing this flowsheet and conceptual layout is itemized in Section 4.3. The envisaged process would lead to a net uranium recovery of approximately 82% with an overall solution concentration ratio of 4.7 to 1 based on the uranium concentration in the combined discharge to the mine sump relative to the uranium concentration in the bioleach feed.

The main process assumption for biomass adsorption in this study, based on currently known technology, is the use of stirred reactors and conventional dewatering/clarification equipment to effect uranium recovery and biomass separation for recycle. Adsorption in columns with pelletized biomass, similar to ion exchange, has not been considered currently feasible due to process unknowns for this type of material at the present time. The process envisaged would be a continuous adsorption/elution process for production of concentrated eluate which would be pumped to surface for eventual uranium recovery in the mill circuit.

Bioleach solution would be collected in a feed tank and pumped to two pH adjustment tanks in series where the pH would be raised with lime from its nominal value of 2.0 to a control value of 3.5. This pH adjustment step is necessary for optimizing the loading characteristics of the biomass. The resulting gypsum/iron precipitate would be removed in a clarifier and pumped as a 25% solids sludge to the main mine sump. The amount of precipitate in this sludge would not be significant compared to the general minewater flows from the mine and hence the precipitate could be pumped to surface where the contained uranium would be recovered.

The clarified solution would then gravitate to a mechanically-agitated bioadsoption tank where recycled and fresh biomass (Rhizopus arrhizus) would be added for uranium adsorption. Biomass make-up, and removal,

would be sufficient to allow for 10 cycles of re-use for the biomass prior to being discarded. The solution/biomass mixture would gravitate to a second clarifier where the biomass would be collected as a dilute underflow and the clarified barren solution overflow would be collected in a tank and recycled to the underground leaching areas. A net minewater make-up of 9.7 m^3/h would be required to balance the solution flow in the bio-leaching blocks. A portion of the biomass clarifier underflow would be discharged directly to the mine sump to effect the 10% biomass bleed.

The recycled barren solution, at pH 3.5 and of low buffering capacity, would be discharged to a mined-out stope to provide some retention time for the pH to drop prior to recycling to the leaching blocks. This is a suitable assumption based on current operating techniques for underground leaching(2). Since the solution would be of low buffering capacity the pyrite reaction products from the in-situ bacterial oxidation conditions would result in a decrease in solution pH to a value approaching 2 which would be compatible with the solution pH in the leaching blocks. At the same time this would lead to replacement of the iron which was precipitated from solution in the pH adjustment step. This process assumption was felt reasonable for purposes of this study especially since any chemical addition step for replacing precipitated iron would only have led to further increases in costs for the Biomass Adsorption Option.

The biomass clarifier underflow would be further dewatered in a centrifuge and the biomass pulp, at approximately 15% solids, would be discharged to an elution tank. The centrate solution (barren) from this centrifuge would be recycled to the biomass clarifier via the bioadsorption tank. The bioadsorption efficiency has been calculated at approximately 78% based on solution analyses and circuit material balances.

The loaded, dewatered, biomass would be eluted with 16.8 g/L $NaHCO_3$ solution for a total retention time of 15 minutes. The calculated elution efficiency of approximately 87% would result in a desorbed

biomass assaying 3.2 kg/t U_3O_8 . The desorbed biomass would be dewatered to 15% solids in a centrifuge and recycled to the bioadsorption tank for re-use. The concentrated eluate collected as centrifuge centrate would discharge by gravity to the main mine sump area along with the 10% biomass bleed and the gypsum/iron precipitate. This solution would then be pumped with minewater to the surface mill facility for uranium recovery.

Reagent requirements for this conceptual process include lime, sodium bicarbonate, flocculant and the Rhizopus arrhizus biomass.

Lime would be delivered to the minesite in bulk and piped underground to a storage silo. It would be fed directly to the pH adjustment tanks by a variable speed screw conveyor at a rate of 3,050 kg/day or 14.8 kg/kg U_3O_8 recovered (95% CaO basis).

Sodium bicarbonate would also be received in bulk and stored underground in a silo. It would be made up as a 16.8 g/L $NaHCO_3$ solution with barren recycle water prior to addition to the elution tank. Consumption has been estimated as 2,660 kg/day or the equivalent of 12.9 kg/kg U_3O_8 recovered.

Flocculant would be received in bags and made up as a 0.05% solution for feeding to the clarifiers. Requirements have been estimated as 7.7 kg/day or 0.04 kg/kg U_3O_8 recovered.

The biomass, Rhizopus arrhizus, is assumed to be manufactured off-site by an independent supplier probably in an air-lift fermentor using established fermentation technology. For the purpose of this study the biomass is assumed to be received in a dry form and transferred to a storage bin underground. It would be added directly to the bioadsorption tank by the use of a variable speed screw conveyor. For a continuous operation with a 10% bleed of spent biomass net consumptions have been estimated at 1,200 kg/day or an equivalent 5.8 kg/kg U_3O_8 recovered.

4.3 DESIGN CRITERIA

The design used for the biomass adsorption flowsheet and associated conceptual layout was obtained from the information generated in the literature review as well as file information available from the project team. For the most part it is based on recent work in this area by active participants(32,38,39,40,41). The criteria details and assumptions used for the Biomass Adsorption Option are detailed below.

4.3.1 General

Bioleach solution flow (m^3/h) - average	- 70
- maximum	- 100
Bioleach solution feed grade (g/L U_3O_8)	- 0.15
Net bioadsorption uranium recovery (%)	- 81.7
Net Concentration Ratio	- 4.7:1
Uranium recovery - kg/day U_3O_8	- 206
- kg/year U_3O_8	- 72,100
Operating days per year	- 350
Operating system	- continuous bioadsortion elution

4.3.2 pH Adjustment

Feed tank retention time (min)	- 5
pH adjustment retention time (min)	- 30
pH adjustment tanks	- 2
Target pH	- 3.5
Lime consumption (95% CaO) - g/L	- 1.7
- kg/day	- 3,050
Lime addition	- dry basis
Gypsum/iron precipitate (t/day)	- 5
Uranium to gypsum/iron precipitate	- 5% of feed
Urannium to mine sump with gypsum/iron precipitate (kg U_3O_8/day)	- 14
Clarifier feed TSS (mg/L)	- 3,000

Clarifier unit area ($m^3/h/m^2$)	- 3
Clarifier net feed rate (m^3/h)	- 70.1
Clarifier underflow density (% solids)	- 25
Clarifier overflow TSS (mg/L)	- 20
Clarifier underflow recycle (%)	- 10
Sludge tank retention time (min)	- 60
Flocculant dosage (mg/L)	- 2

4.3.3 Bioadsorption

Retention time (min)	- 30
Biomass type	- Rhizopus arrhizus
Biomass bleed - %	- 10
Biomass make-up - kg/day (dry)	- 1,200
Biomass specific gravity (t/m^3)	- 1.0
Biomass bulk density (t/m^3)	- 0.2
Biomass loading - g/kg U_3O_8(dry basis)	- 25
Number of cycles for biomass	- 10
Bioadsorption efficiency (%, based on solution analysis)	- 77.6
Biomass clarifier feed TSS (mg/L)	- 5,000
Biomass clarifier unit area ($m^3/h/m^2$)	- 2
Biomass clarifier net feed rate (m^3/h)	- 89.5
Biomass clarifier underflow density (% solids)	- 2
Biomass clarifier underflow bleed (m^3/h)	- 2.6
Biomass clarifier overflow TSS	- trace
Biomass clarifier overflow analysis (g/L U_3O_8)	- 0.032
Biomass clarifier underflow recycle (%)	- 10
Flocculant dosage (mg/L)	- 2
Barren recycle tank retention time (min)	- 5
Net barren recycle flow (m^3/h)	- 60.3
Uranium to mine sump with biomass bleed (kg/day U_3O_8)	- 32

4.3.4 Elution

Loaded biomass centrifuge feed	- m^3/h	- 20.0
	- t/h solids	- 0.40
	- % solids	- 2
Loaded biomass centrifuge centrate	- m /h	- 17.3
	- TSS	- trace
	- g/L U_3O_8	- 0.032
Loaded biomass centrifuge sludge	- t/h	- 0.40
	- % solids	- 15
	- g/kg U_3O_8	- 25
Eluant type		- $NaHCO_3$
Eluant concentration (g/L $NaHCO_3$)		- 16.8
Eluant flow (m^3/h)		- 6.6
Sodium bicarbonate consumption (kg/day)		- 2,660
Elution retention time (min)		- 15
Desorbed biomass centrifuge feed	- m^3/h	- 9.3
	- t/h solids	- 0.40
	- % solids	- 4.3
Desorbed biomass centrifuge centrate	- m^3/h	- 6.6
	- TSS	- trace
Concentrated eluate analysis (g/L U_3O_8)		- 1.01
Desorbed biomass centrifuge sludge	- t/h	- 0.40
	- % solids	- 15
Desorbed biomass analysis - g/kg U_3O_8		- 3.2
Eluate tank retention time (min)		- 5
Elution efficiency (%)		- 87.2
Elution concentration ratio		- 6.7:1
Uranium to mine sump with concentrated eluate (kg U_3O_8/day)		- 160
Total uranium to mine sump (kg U_3O_8/day)		- 206
Net concentration ratio		- 4.7:1

4.3.5 Reagents

Hydrated lime:	consumption (95% CaO)- kg/day	- 3,050
	- kg/kg U_3O_8	- 14.8
	delivery	- bulk
	storage (days)	- 3
	strength	- direct
Sodium bicarbonate:	consumption (100% $NaHCO_3$)	
	- kg/day	-2,660
	- kg/kg U_3O_8	-12.9
	delivery	- bulk
	storage (days)	- 4.5
	strength (g/L $NaHCO_3$)	- 16.8
Flocculant:	consumption - kg/day	- 7.7
	- kg/kg U_3O_8	- 0.04
	delivery	- bags
	strength (%)	- 0.05
Biomass:	type	- Rhizopus arrhizus
	production method	- air-lift fermentor
	supplier	- off-site
	consumption - kg/day	- 1,200
	- kg/kg U_3O_8	- 5.8
	delivery	- bulk
	storage (days)	- 3
	strength	- direct

4.4 CAPITAL COST ESTIMATE

4.4.1 Summary

The cost summary presented in Table 4-1 provides the estimated capital costs of the Biomass Adsorption Option for recovering uranium from underground bioleach solution with a solution feed rate of 70 m^3/h. These estimated costs total $3,736,600 in first quarter 1986 Canadian dollars and provide for all costs of the biomass adsorption facility.

The estimate includes:

- underground excavation and concrete pad
- complete biomass adsorption circuit including reagent storage underground
- indirect costs of construction
- engineering, procurement, and construction management
- capital spares inventory which includes initial fills
- contingency allowance.

The capital cost estimate details for the Biomass Adsorption Option are provided in Appendix B.

The total unit capital costs of this option can be interpreted as follows:

- $6.35 per annual cubic meter of bioleach solution processed
- $51.83 per annual kilogram of U_3O_8 recovered
- $23.51 per annual pound of U_3O_8 recovered.

These cost estimates do not include any provision for pre-production construction interest as the total value of this interest is dependent on the method of financing and the equity or credit position of the specific investor. These pre-production interest costs can have a significant impact on the total costs of any project and this should be kept in mind when reviewing the cost estimates given in this report.

TABLE 4-1

BIOMASS ADSORPTION OPTION
CAPITAL COST ESTIMATE

Direct Costs	$
Underground Excavation	334,720
Equipment	1,227,651
Mechanical	303,975
Piping	337,000
Electrical	464,045
Instrumentation	122,500
Total Direct Costs	2,789,891
Indirect Costs	
Construction Indirects (8%)	223,191
Engineering,Procurement and Construction Management (12%)	334,787
Capital Spares	49,031
Sub-Total	3,396,900
Contingency (10%)	339,690
TOTAL ESTIMATED CAPITAL COSTS	3,736,590
	say, 3,736,600

4.4.2 Basis of Estimate

The capital cost estimate is based on the following information:

- flowsheet and conceptual layout prepared by Melis Consulting Engineers Ltd.
- budget quotations on major equipment
- pricing data from Melis Consulting Engineers Ltd. and Kilborn Ltd.'s files and records to arrive at factors for mechanical, piping, electrical and instrumentation costs. These factors have been adjusted to reflect the degree of complexity of the proposed process and the relative cost of unit equipment.

For purposes of this study it has been assumed that:

- all construction services would be tendered by fixed price contract
- the average labour rate is $35.00 per hour based on a 40-hour week. This labour rate includes provision for direct wages, payroll burdens and benefits, small tools and consumables, site preparation and the contractor's overhead and fee
- construction labour force is available in the immediate surrounding area
- the cumulative price factor applied to base prices for process equipment is eighteen percent,which includes:
  - provincial sales tax - 7%
  - freight allowance - 5%
  - miscellaneous minor materials allowance - 5%
- all costs are reported in first quarter 1986 Canadian dollars.

No allowances have been made in the capital estimate for:

- escalation beyond the first quarter 1986
- pre-production construction interest
- working capital
- production equipment or materials
- start-up assistance

- cost of pre-appropriation studies and development work
- owners' costs such as owners' representative
- general property facilities such as administration offices
- pumping bioleach solutions from the underground leaching areas to the treatment facility
- pumping of the concentrated solution to surface with the general minewater
- decommissioning.

4.4.3 Methodology

Equipment prices were obtained on an individual basis from budget quotations and historical data. Mechanical installation costs, including field material, were factored from the equipment costs. Piping, electrical and instrumentation costs were factored from the total installed mechanical costs. Building structure costs for each area were based on volumes and volume unit prices derived from historical data. The factors used in deriving the total installed mechanical costs have been adjusted to reflect the complexity and relative cost of unit equipment for the proposed process.

The allowance for construction indirects has been estimated as eight percent of the total direct costs. It includes:

- temporary services
- permits, insurance and bonds
- rental of mobile equipment
- on-site construction supervision facility
- pre-construction geotechnical investigations.

The engineering, procurement and construction management function has been estimated as twelve percent of the direct costs. As well as design engineering, this includes provision for project management, cost control, scheduling, contract administration and inspection services.

An allowance of five percent of home office costs has been included for capital spares inventory which would include the cost of initial fills in the circuit.

The contingency allowance provides for additions and unidentified items within the stated scope of work but it is not intended to cover major changes in scope.

4.5 OPERATING COST ESTIMATE

4.5.1 Summary

The biomass adsorption process operating costs, based on the design basis detailed in Section 4.3, total $3,519,820 annually in first quarter 1986 Canadian dollars, including a five percent contingency allowance. This annual operating cost is equivalent to $6.00/m^3 of bioleach solution processed or $48.83/kg U_3O_8 ($22.13/lb U_3O_8) based on a total of 72,100 kg U_3O_8 (158,951 lbs U_3O_8) of recovered uranium per year for the Biomass Adsorption Option.

A summary of the operating costs for a 350 day per year operation is presented in Table 4-2 with cost details given in Tables 4-3 and 4-4. The operating costs include all costs associated with the recovery of uranium from underground bioleach solutions by the proposed biomass adsorption process. The processing costs extend from receipt of the bioleach solution underground to the production of concentrated solution discharged to the main mine sump. These costs include manpower, reagents and supplies, power, and the cost of analysis for circuit operation. It does not include any costs associated with the actual underground leaching nor does it include any costs associated with downstream processing of the uranium in the surface milling circuits.

4.5.2 Manpower

The annual manpower costs detailed in Table 4-3 include the manpower estimated to be required for the operation of the biomass adsorption circuit underground. The average annual total manpower is estimated at eight men for meeting operation and maintenance requirements. A twenty-five percent payroll burden has been assumed in addition to the indicated salary rates.

TABLE 4-2

BIOMASS ADSORPTION OPTION
ANNUAL OPERATING COSTS

Item	Annual $	$/m^3	$/kg U_3O_8	$/lb U_3O_8
Manpower	345,300	0.59	4.79	2.17
Reagents and Supplies	2,925,310	4.97	40.57	18.40
Power	51,600	0.09	0.72	0.32
Analysis	30,000	0.05	0.42	0.19
Sub-Total	3,352,210	5.70	46.50	21.08
Contingency (5%)	167,610	0.30	2.33	1.05
TOTAL	3,519,820	6.00	48.83	22.13

TABLE 4-3

BIOMASS ADSORPTION OPTION
ANNUAL MANPOWER COSTS

Employee	Requirements	Number	Rate	Annual $
Supervision-Operation	0.75 man/day	0.75	$3,500/month	31,500
-Maintenance	0.25 man/day	0.25	$3,500/month	10,500
Operator	1.5 men/shift	6.0	$16.00/hour	201,600
Maintenance	20 mandays/mth	1.0	$17.00/hour	32,640
Sub-Total		8.0		276,240
Payroll Burden (25%)				69,060
TOTAL ESTIMATED MANPOWER COSTS				345,300

TABLE 4-4

BIOMASS ADSORPTION OPTION
ANNUAL REAGENTS AND SUPPLIES COSTS

Item	Unit Cost ($/kg)	Consumption		Annual $
		kg/kg U_3O_8 (1)	kg/year	
Lime (95% CaO)	0.083	14.8	1,067,080	88,568
Sodium Bicarbonate	0.84	12.9	930,090	781,280
Flocculant	3.00	0.04	2,884	8,652
Biomass	4.50	5.8	418,180	1,881,810
Maintenance Supplies				100,000
Operating Supplies				60,000
Laboratory Supplies				5,000
TOTAL				2,925,310

Notes: (1) Reagent consumptions are based on 72,100 kg of recovered uranium (U_3O_8) per year.

The supervision allowance for operations is based on a supervisor spending seventy-five percent of his time on dayshift monitoring the operation of the biomass adsorption circuit. Maintenance supervision would be less with a supervisor spending approximately twenty-five percent of his time in the area.

Operating manpower for the underground biomass adsorption circuit is based on a requirement of one and a half men per shift three shifts per day, 350 days per year. A maintenance allowance of 20 mandays per month has been assumed for maintenance of the equipment in the biomass adsorption circuit.

4.5.3 Reagents and Supplies

The cost of reagents and supplies detailed in Table 4-4 is based on consumption rates given in the biomass adsorption design basis and the delivered price of reagents to the facility. The annual consumption is based on 72,100 kg of recovered uranium (U_3O_8) per year.

The lime consumption of 14.8 kg/kg U_3O_8 (95% CaO basis) is based on a 1.7 g/L addition rate to the bioleach solution for raising the pH from 2.0 to 3.5. The sodium bicarbonate requirement is based on eluting the loaded biomass with 16.8 g/L $NaHCO_3$ solution. The flocculant consumption of 0.04 kg/kg U_3O_8 includes the requirement for both the gypsum/iron clarifier and the biomass clarifier. The net biomass addition of 5.8 kg/kg U_3O_8 is based on the need for a 10% bleed in order to maintain adequate adsorption capacity in the biomass. The biomass unit cost of $4.50/kg is an estimated production cost of biomass using established fermentation technology(40).

Lump sum estimates have been included for laboratory, maintenance and operating supplies sufficient to meet the requirement for operation of the biomass adsorption circuit.

4.5.4. Power

The total power demand for operating the biomass adsorption circuit is estimated at approximately 260 kW. The total annual power cost is thus estimated at $51,600 assuming an 85% power efficiency factor and a power cost of $0.02/kWh for a 350 day per year operation.

4.5.5 Analysis

The lump sum estimate of $30,000 for analytical costs, sufficient for 3,000 analyses per year, would meet the requirements for daily metallurgical control of the various unit operations of the biomass adsorption circuit.

4.6 TECHNICAL EVALUATION

4.6.1 Discussion

As identified in the literature review the capability of microorganisms to extract metal ions from aqueous solutions is well known with recent studies having shown that microorganisms act as a natural sink for uranium which in turn can result in a concentration effect for dilute solutions.

Studies specifically related to the potential of uranium concentration and recovery by biomass from bioleach solutions have recently been accomplished (32,38,39,40,41). The results identified in these reports have been used as the basis for the biomass adsorption recovery process described in this study.

Based on currently known technology the use of stirred reactors and conventional dewatering/clarification equipment to effect uranium recovery and biomass separation for recycle is the only practical processing option which can be assumed for the study (42,43). The use of biomass in a pelletized form is claimed to be possible (41,44) but process unknowns for this type of material will require more investigations to prove its feasibility (42,43).

The use of stirred reactors as opposed to large diameter columns for contacting biomass with bioleach solution has been selected as the most reliable method given the unknowns of physical integrity and flow-through characteristics of biomass. Clarifiers and centrifuge have been selected for liquid/solid separation requirements of the biomass adsorption process since these are common equipment for handling similar forms of materials (gelatinous precipitates etc.). Unit area requirements for clarifiers are considered reasonable in terms of space utilization. Typically clarifiers are provided with a deep centre well which results in a thick bed of suspended solids to

filter upflowing solution. Centrifuges would provide a means of maximizing the density of the biomass for recycle to the biomass adsorption circuit.

The use of biomass necessitates a pH adjustment step to increase the bioleach solution pH to a range which allows a reasonable uranium loading on the biomass (38,39). A value of 3.5 has been chosen as the target pH to minimize lime consumption and to minimize acidity reductions in the barren recycle solution (37,39). If the control pH approached a value of 4.0 it would lead to an increased loss of iron and calcium sulphate from solution by precipitation thus reducing the bioleaching efficiency of the recycled barren solution. A pH 3.5 level would also minimize the amount of uranium which would co-precipitate with the gypsum/iron precipitate.

A retention time of 30 minutes for this pH adjustment step should be adequate based on other studies (45). The lime consumption figures and precipitate estimates are based on the chemical make-up of the bioleachate solution (42). These values are generally the same order-of-magnitude as other values generated in metallurgical testwork on uranium ores (45). The gypsum/iron precipitate clarifier design values and flocculant dosage are based on actual values generated during testwork on similar solutions (45).

The bioadsorption design details assumed in the study are based on available information from recent test results (38,39,40,41,42,44). The biomass type, Rhizopus arrhizus, has been determined to be the most suited for this application (38) and was selected as the biomass of choice for the study. The specific gravity of biomass is close to unity (44) and has been assumed as 1.0 for this study. Test results (38) have indicated a bulk density of 0.2 which was used for sizing storage requirements.

A biomass loading of 25 g U_3O_8 /kg is a realistic assumption for the loading characteristics of biomass when processing pH 3.5 bioleach solutions (40,42). The net bioadsorption efficiency of 77.6% is based on a feed solution containing 0.143 g/L U_3O_8 and a barren solution of 0.032 g/L U_3O_8(42). A 10% biomass bleed has been assumed based on test results which showed the biomass to be suitable for 10 adsorption/desorption cycles before requiring replacement (38).

The biomass clarifier and centrifuge operating conditions are based on design criteria which are believed to be realistic for the biomass based on values of products with similar characteristics (45,46,47). The use of flocculant has been assumed as necessary to assist in the densifying of the biomass sludge prior to centrifuging.

In earlier studies elution of loaded biomass had been shown to be efficient when using 0.1 M $NaHCO_3$ solution(38,40) but a 0.2 M solution has been assumed in this study since more recent work has indicated that this would be a more efficient concentration for elution of loaded biomass(42). In this study, based on uranium balances, the calculated elution efficiency is approximately 87% and the concentrated eluate analysis is 1.09 g/L U_3O_8. The analysis of the desorbed biomass, which is recycled to the bioadsorption tank, is 3.2 g/kg U_3O_8.

Based on a previous evaluation (40) the biomass, Rhizopus arrhizus, is assumed to be produced off-site by an independent supplier in an air-lift fermentor using established fermentation technology.

4.6.2 Recommendations

The hydrometallurgical process envisaged for the Biomass Adsorption Option is based on current technology and as such cannot really be considered as practical nor suitable for an in-situ application in an underground environment in light of technical advances which need to be made for this option.

The high capital cost inherent with the Biomass Adsorption Option are due to unknowns regarding its physical integrity .These unknowns led to the selection of equipment which was based on current technology for biomass adsorption, hence the type of equipment selected is of necessity more costly than equipment for a column type operation (such as the ion exchange process). A pelletized form of immobilized biomass would need to be available to make a column type operation viable for biomass adsorption in order to arrive at lower capital costs.

The requirement for a pH adjustment step for the bioleach solution ahead of biomass adsorption, and the requirement for elution of loaded biomass to allow recycle of biomass to the bioadsorption circuit also contribute significantly to the high capital costs of this option. The pH adjustment circuit represents approximately 23% of the total capital costs for this option while the elution circuit is approximately 19% of the total capital costs for biomass adsorption. Hence capital cost reductions could be realized if an organism could be found for processing low pH solutions thereby eliminating the pH adjustment step and if biomass production costs could be decreased to permit single usage of biomass thereby eliminating any requirement for recycling biomass which in turn would eliminate the need for an elution circuit.

The significantly high operating costs which have been identified for the Biomass Adsorption Option are generally due to the following:

- poor uranium loading efficiencies (only ten to twenty percent of resin capacities) causing high biomass addition rates to the adsorption circuit
- high biomass replacement costs since the biomass is only efficient for approximately ten cycles
- high production costs for biomass
- greater manpower requirements due to the relative complexity of the circuit
- high consumption, and hence cost, of the sodium bicarbonate eluting agent.

It has been estimated that the elution circuit represents approximately 28% of the annual operating costs for biomass adsorption while the pH

adjustment circuit only represents about 8% of the annual operating costs.

Since the use of biomass to recover uranium from bioleach solutions is a relatively new concept additional research is required to more adequately define the proposed process and to arrive at improvements for hydrometallurgical applications. The following items are therefore suggested avenues of future experimental work to render biomass adsorption a more viable option for uranium recovery from bioleach solutions:

- investigate immobilization techniques for cultures that would allow the retention of high bioadsorption capacity while maintaining high physical strength during mechanical and chemical processing
- test the physical strength and efficiency of a pelletized form of biomass which is claimed to be available (38,41) and determine its applicability for a column-type operation
- assess other possible lower cost solution pre-treatment for Rhizopus arrhizus cultures which would lead to improved loading efficiencies
- investigate alternate cultures which would not require pre-treatment such as pH adjustment but yield equivalent adsorption capacities
- determine the limits of the degree of biomass recycle that is achievable to better define biomass make-up rates
- investigate alternate eluant chemicals which may be better suited for underground operations
- identify the structural groups responsible for sorption thereby potentially leading to enhanced bioadsorption capacities or cell-free media
- determine actual reagent consumptions for the process conditions assumed in the study
- determine the effect, if any, of the eluant, $NaHCO_3$, on uranium extractions in underground bioleaching
- quantify the CO_2 gas generation when utilizing $NaHCO_3$ for

loaded biomass desorption in terms of potential concerns in an underground environment

- measure the specific gravity and bulk density of biomass and assess its handling characteristics
- investigate methods of producing biomass to reduce unit costs of the product and hence eliminate the need for an elution circuit
- confirm the physical characteristics of biomass for equipment sizing
- carry out appropriate tests, preferably in a small-scale continuous mode, to confirm the operating criteria assumed in the study.

5.0 REVERSE OSMOSIS OPTION

5.1 LITERATURE REVIEW

5.1.1 Background

Reverse osmosis (R.O.) is a pressure driven membrane separation process. High separation of solutes from water can be obtained due to the selective permeability characteristics of reverse osmosis polymeric membranes. The process requires no addition of heat or other components and can be run on a continuous basis with little maintenance other than membrane replacement.

The principle of R.O. is that pure water is forced through a membrane under pressure sufficient to overcome osmotic pressure (permeate) while dissolved salts are prevented from passing through the membrane and remain in the concentrate. R.O. is a proven process for upgrading brackish water for potable use where the permeate is the product. In this study the desired product would be the concentrate since it would be upgraded in uranium concentration.

5.1.2 Details

There is little literature available that applies directly to the problem of recovering uranium from bioleach solutions using reverse osmosis. A limited bench scale study completed by Zenon Environmental Inc. for Denison Mines Limited is probably the most relevant source of information(39). A Film Tec FT-30 spiral wound membrane module was used to treat leachate solution provided by Denison Mines Limited. Good flux rates (0.6-0.7 $m^3/m^2/d$) and greater than 99 percent uranium separation were observed.

These data are encouraging and useful for preliminary scale-up calculations but several issues require further analysis. The bioleach

solution tested contained a higher concentration of iron but lower concentrations of silica and sulphate than the design bioleach characteristics assumed in this study. The Zenon testing was done as a batch process with the concentrate recycled to the feed tank. This simulates the gradual increase in concentrate strength that occurs between the inlet and outlet of a series of R.O. modules but it provides no information on the effect of time on system performance. The bench apparatus was shut down once 80 percent recovery was achieved (ie., concentrate volume reduced to 20 percent of initial feed volume).

A patent issued to Cleary and Stana in 1980 dealt with the recovery of uranium from solutions containing other inorganic salts using cellulose acetate membranes (48). Good fluxes were achieved with 91 percent or greater separation of uranium but cellulose acetate membranes would not be appropriate for the present study since these membranes cannot be operated at low pH.

A second patent issued to Stana and Tiepel discusses a scheme for recovering uranium in-situ from leachate solutions(49). It was suggested that ultrafiltration would be able to recover uranium providing that an appropriate uranium complexing agent was added. The process differs from reverse osmosis in that most of the inorganic salts present would pass through the membrane. A pretreatment system to remove calcium salts by raising the pH and filtering out the precipitated salts was recommended.

Several papers(50,51,52) have been published on the general problem of removing uranium from low level radioactive water and wastewater including radioactive laundry water. Some of these references are useful to illustrate the feasibility of separating uranium from water using reverse osmosis. Most notable of these is a series of articles by Roberts and various coworkers published as a series of reports to the US Department of Energy(50). It was found that uranium was rejected at low concentrations by the ultrafiltration membranes used. Other studies on laundry water have found high separation when reverse osmosis type membranes were used(51). Typically these studies used

cellulose acetate membranes at close to neutral pH with 99% or greater uranium separation being observed. Similar results were obtained on synthetic feeds under similar operating conditions in other studies by Sastri and Ashbrook(52).

5.1.3 Technical Considerations

The operating conditions for a reverse osmosis system are specified by the composition, flowrate, temperature, and pH of the feed stream. In addition it is also necessary to specify the operating pressure and ultimately to specify the exact membrane modules to be employed in order to complete sizing and economic evaluations of a reverse osmosis plant.

Feed Composition

The concentration and chemical composition of dissolved solids determines the osmotic pressure of the feed solution. Osmotic pressure is important in determining the operating pressure and flux rate for R.O. membranes. Since the total dissolved solids content is low (2700-3500 mg/L) the osmotic pressure should also be low, about 150-200 kPa based on equivalent NaCl concentration(56). For the desired product recovery of the Reverse Osmosis Option the osmotic pressure is negligible and can therefore be safely ignored as a first approximation.

The presence of calcium ions may cause fouling of the membranes. However, $CaSO_4$, which is present in this system, is much less a problem than the less soluble $CaCO_3$.

Feed Flowrate

The feed rate coupled with the desired (or obtainable) recovery essentially fixes the amount of membrane area that is required. In addition, the feed rate will determine the linear velocity of the high

pressure solution over the surface of the membrane. This velocity in turn determines the concentration polarization in the R.O. module. A membrane manufacturer would specify the velocity required to minimize concentration polarization effects.

An R.O. system should be designed to handle the peak sustained flowrate of 100 m^3 /h and still operate at the design flowrate of 70 m^3/h. Typically, reverse osmosis systems have good turndown characteristics so this presents no special problem.

Feed Temperature

It is normally advantageous to operate at higher temperatures since permeate flux increases with temperature(56). However, the low solution temperatures in this application are actually preferred since the solubility of calcium salts would be higher at lower temperature, thus decreasing fouling, and the chemical degradation of the membrane by acid hydrolysis should be less at lower temperature, thus increasing membrane life. Separation would only be marginally lower at low temperatures.

Feed pH

Some manufacturers advertise membranes which can be operated continuously at pH 2 and which can stand occasional short term exposure to pH 1.5. Generally speaking, it will be important to maintain the pH of the bioleach solution at no less than pH 2 as there are few if any membranes which would provide satisfactory services at lower pH. Most cellulosic membrane materials cannot be operated at a pH 2.7 or lower hence these types of membranes are specifically not recommended(57).

Operating Pressure

Permeate flux is approximately linear with pressure and separation

increases with increasing pressure so that a high operating pressure is desirable. On the other hand very high pressures lead to higher costs due to the need for more expensive pumps(58). An operating pressure of 2,000 kPa with a "brackish water" type membrane unit is therefore recommended with the design pressure being 2,750 kPa in order to increase flux rates by increasing pressure if necessary. This pressure can be considered as economically suitable in terms of the flux rates which can be obtained and thus operating pressures any higher than this are not really warranted.

Membrane Type

Tubular and spiral wound are the standard membrane configurations commercially available. Tubular membranes can handle suspended solids better but are more costly per unit area of membrane surface than spiral wound membranes(59). There is probably a larger choice of spiral wound membranes that would be able to operate at the low design pH. Hollow fibre is a third class of membrane but it cannot be used when there is any suspended material or any chance of membrane fouling. This type would therefore not be suitable for the bioleach solution. The exact choice of a membrane module would depend on a detailed economic evaluation but for the purpose of this study a spiral wound membrane can be assumed due to its better compatibility with lower pH solutions.

Feed Pretreatment

The expected concentration of suspended solids in the bioleach solution feed is too high for spiral wound type membrane systems, hence a solid removal pretreatment process would be necessary. This would include a cartridge or bag type prefiltration system immediately ahead of the R.O. unit to guarantee against membrane damage or fouling.

Recovery

The limited solubility of $CaSO_4$ in water will probably limit, and hence dictate, the recovery that can be obtained in a reverse osmosis process. The limit of scaling can be extended somewhat by the addition of an antiscalant. A rough relationship between CaSO concentration and the recovery factor (at 25°C) with antiscalant present is given as:

$$\text{mg/L Ca} \times \text{mg/L } SO_4 \times (1.5/1\text{-}R))^2 < 1.3 \times 10^7 \quad (60)$$

where R is the recovery factor defined as permeate flowrate/feed flow rate. Using the design feed concentrations of Ca and SO_4 the maximum recovery factor can be estimated from this equation as R=0.70. An R value of 0.70 is thus the maximum which could be anticipated based on calcium sulphate solubility. The above equation is for room temperature so this estimate can be considered conservative.

If the R value is 0.70 then the volume recovery to the permeate (barren recycle solution) wold be 70% and the volume recovery to the concentrated solution would be 30%. Assuming that the membrane completely rejects uranium to the concentrated solution(39) then the corresponding increase in uranium concentration would be 3.3 times the concentration in the bioleach feed.

The peak concentration of 15,000 mg/L SO_4, the sulphate level assumed for the study, is very high and would almost certainly foul the membrane based on solubility considerations. Some of the dissolved solids will be rejected by the membrane which should prevent gradual accumulation of sulphate and other ions in the circuit as permeate is recycled to the block area for re-use. This will help control the maximum sulphate concentration to relatively low values.

The presence of silica in the feed may also cause solubility problems since the solubility of silica in water is only about 150 mg/L (at 25°C). Again this suggests that no more than 70% recovery should be attempted for continuous operation to avoid silica saturation in the concentrate.

Membrane Cleaning

It is almost always necessary to include some sort of membrane cleaning cycle in a reverse osmosis process, particularly with applications such as this where membrane fouling may be a problem. Cleaning the membrane for a short period of time on a regular schedule would greatly extend membrane life. The exact nature of the cleaning step would have to be determined experimentally but typically sodium hydrosulphite solution has been used for iron scaling and ammonium bifluoride has been used for silica scaling(60).

5.2 GENERAL PROCESS DESCRIPTION

The detailed flowsheet and material balance for a conceptual uranium recovery circuit from bioleach solution using reverse osmosis is presented on Drawing No. 100-10-3. A conceptual layout of the equipment is shown on Drawing No. 100-20-3. The net uranium recovery for this process has been estimated at 99 % with a solution concentration ratio of 3.3 to 1 based on the uranium concentration in the retenate (the concentrated solution) relative to the feed concentration in the bioleach solution. As discussed above this concentration ratio is limited by the solubility of calcium sulphate present in the solution and is also based on the rejection of heavy metal ions(52,60).

In this process the pre-treatment requirement ahead of reverse osmosis is clarification of the bioleach solution to remove suspended particles of 5 microns or less for preventing fouling of the reverse osmosis membrane(54). Bioleach solution would be collected in a tank and processed through sand filters to remove the majority of suspended solids.

The flow would be split to two dual-media sand filters in parallel to accomodate the total flow of 70 m^3/h. A third sand filter would be needed for standby or for use during the backwash mode. Sand filter discharge would be retained in a storage tank prior to further clarification. Backwash water, containing the suspended solids, would be discharged to the mine sump for disposal. It has been assumed that each filter would need to be backwashed every 8 hours.

From operating experience clarities of 1 mg/L TSS in the sand filter discharge, from a feed of 10 mg/L TSS, should be possible with efficient operation of sand filters. In order to accomodate any upset conditions however, and to remove particles down to the 5-micron size, a bag filter has been included in the process flow for further clarification of the sand filter discharge ahead of reverse osmosis. The bag filter discharge, clarified bioleach solution, would be stored in a

tank for feeding to the reverse osmosis unit. It has been assumed that spent bags in the bag filter unit, containing an estimated 100 kg of solids, would be replaced every two months.

Anti-scaling agent, suitable for minimizing gypsum-type scaling, would be added to the clarified bioleach solution prior to feeding to the reverse osmosis unit. A multi-stage (five) pump would be used to achieve the high feed pressure (2,000 kPa) requirements of reverse osmosis. Retenate, or upgraded concentrate solution, would be discharged to a tank through a pressure-relief valve and report by gravity to the mine sump for pumping to the surface mill facility for uranium recovery. Permeate would be collected in a tank and the majority would be recycled to the underground leaching areas. Approximately 20% of the permeate would be used for backwashing the sand filters. The net minewater make-up to the bio-leaching blocks would therefore be 30 m^3/h. Since the reverse osmosis unit would remove the majority of the dissolved solids from the permeate it would be necessary to store the permeate in a mined-out stope to allow the nutrients to re-establish themselves in solution prior to recycling the permeate to the bio-leaching blocks. This aspect would need to be investigated in detail in order to identify the extent of nutrient removal by a reverse osmosis process.

The net uranium recovery efficiency for the reverse osmosis process would be expected to be 99% or better(39). The anticipated permeate/feed ratio of 0.7 to 1 would lead to a concentration ratio of 3.3 to 1 in terms of uranium solution analysis.

The type of reverse osmosis unit assumed for the study is spiral wound modules equipped with thin-film membrane material such as Film Tec FT 30. Cellulose acetate membranes would not be suitable due to their susceptibility to acidic solutions (pH 2.7 or lower). Assuming a solution flux rate (specific flowrate) of 27 m^3/module per day a total of 62 modules would be required arranged in a parallel configuration. Since there is no long-term operating data for this type of solution

it has been assumed that membrane replacement would be required every 3 years. The useful life of R.O. membranes is very much dependent on the nature of the feed solution and the operating characteristics of the process.

Iron scaling and silica scaling could be a problem with a reverse osmosis operation on bioleach solutions containing high levels of iron and significant concentrations of silica(60). A cleaning chemical make-up tank, for use in treating the membranes on a regular basis, has been included in the conceptual reverse osmosis process. Sodium hydrosulphite has been assumed as the cleaning agent for removing iron scaling and ammonium bifluoride for removing silica contamination(60). It has been assumed that cleaning of the membranes would be necessary every two weeks in order to maintain high uranium recoveries and adequate flux rates. Spent cleaning solutions would be discharged to the mine sump.

Both sodium hydrosulphite and ammonium bifluoride would be received in bags and made up as 3% solution for feeding to the reverse osmosis modules. Two-stage pumps would be used to achieve moderate feed pressures at the membranes. Requirements for each reagent would be an average of 43 kg/day based on a two-hour bi-weekly application of 3% solution at a flowrate of 10 m^3/h. This is equivalent to a reagent consumption of 0.17 kg/kg U_3O_8 recovered.

Anti-scaling agent would be received in drums and added directly to the reverse osmosis feed tank. At a 10 mg/L dosage rate the consumption would be 16.8 kg/day or 0.07 kg/kg U_3O_8 recovered.

5.3 DESIGN CRITERIA

5.3.1 General

Bioleach solution flow (m^3/h) - average	- 70
- maximum	- 100
Bioleach solution analysis (g/L U_3O_8)	- 0.15
Net reverse osmosis uranium recovery (%)	- 99
Concentration ratio	- 3.3:1
Uranium recovery - kg/day U_3O_8	- 249.5
- kg/year U_3O_8	- 87,325
Operating days per year	- 350
Operating system	- clarification, reverse osmosis

5.3.2 Clarification

Sand filter feed tank capacity (min)	- 5
Sand filter feed rate (m^3/h)	- 70
Sand filter unit flowrate ($m^3/h/m^2$)	- 16
Bed type	- dual media
Media type	- anthracite and sand
Sand filter feed TSS (mg/L)	- 10
Sand filter discharge TSS (mg/L)	- 1
Backwash frequency	- every 8 hours
Backwash time (min)	- 20
Backwash solution	- permeate
Backwash flow (m^3/h)	- 70
Backwash solution discharge point	- mine sump
Bag filter feed tank capacity (min)	- 5
Bag filter size limitation (micron)	- 5
Bag filter unit flowrate ($m^3/h/m^2$)	- 50
Bag filter total filtration area (m^2)	- 2
Bag filter discharge TSS	- nil
Bag change frequency	- bi-monthly
Bag filter solids (kg/month)	- 50

5.3.3 Reverse Osmosis

Reverse osmosis feed tank capacity (min)	- 5
Anti-scaling agent addition rate (mg/L)	- 10
Anti-scaling agent type	- Flocon 100 or equivalent
Anti-scaling agent consumption- kg/day	- 16.8
- kg/kg U_3O_8	- 0.07
Operating pressure (kPa) - average	- 2,000
- design	- 2,750
Operating temperature (^oC)	- 12
Flux rate at 25^{o}C ($m^3/d/m^2$)	- 0.6
Flux rate at 12^{o}C ($m^3/d/m^2$)	- 0.41
Uranium recovery in concentrate (%)	- 99
TDS recovery in concentrate (%)	- 85
Permeate: feed ratio	- 0.70:1
Concentrate: feed ratio	- 0.30:1
Concentration ratio	- 3.3:1
Reverse osmosis membrane type	- Film Tec FT 30
Module - type	- spiral wound
- size	-203 mm x 1016 mm
Module flux rate at 12^{o}C, 2,000 kPa (m^3/d/module)	- 27
Total number of modules	- 62
Module arrangement	- parallel
Permeate recycle tank capacity (min)	- 30
Concentrate tank capacity (min)	- 5
Membrane replacement	- every 3 years

5.3.4 Membrane Cleaning

Reagent type for iron scaling	- $Na_2S_2O_4.2H_2O$
Reagent type for silica scaling	- NH_4HF_2
Delivery	- bags
Solution strength (%)	- 3
Cleaning frequency	- bi-weekly
Cleaning time (hours)	- 2
Flow (m^3/h)	- 10
Average $Na_2S_2O_4.2H_2O$ consumption - kg/day	- 43
- kg/kg U_3O_8	- 0.17
Average NH_4HF_2 consumption - kg/day	- 43
- kg/kg U_3O_8	- 0.17

5.4 CAPITAL COST ESTIMATE

5.4.1 Summary

The cost summary presented in Table 5-1 provides the estimated capital costs of the Reverse Osmosis Option for recovering uranium from underground bioleach solution with a solution feed rate of 70 m^3/h. These estimated costs total $3,680,500 in first quarter 1986 Canadian dollars and provide for all costs of the reverse osmosis facility.

The estimate includes:

- underground excavation and concrete pad
- complete clarification and reverse osmosis circuit
- indirect costs of construction
- engineering, procurement, and construction management
- capital spares inventory which includes initial fills
- contingency allowance.

The capital cost estimate details for the Reverse Osmosis Option are provided in Appendix B.

The total unit capital costs of this option can be interpreted as follows:

- $6.26 per annual cubic meter of bioleach solution processed
- $42.15 per annual kilogram of U_3O_8 recovered
- $19.12 per annual pound of U_3O_8 recovered.

These cost estimates do not include any provision for pre-production construction interest as the total value of this interest is dependent on the method of financing and the equity or credit position of the specific investor. These pre-production interest costs can have a significant impact on the total costs of any project and this should be kept in mind when reviewing the cost estimates given in this report.

TABLE 5-1

REVERSE OSMOSIS OPTION
CAPITAL COST ESTIMATE

Direct Costs	$
Underground Excavation	135,200
Equipment	1,601,087
Mechanical	254,800
Piping	222,700
Electrical	404,050
Instrumentation	111,400
Total Direct Costs	2,729,237
Indirect Costs	
Construction Indirects (8%)	218,339
Engineering,Procurement and Construction Management (12%)	327,508
Capital Spares	70,802
Sub-Total	3,345,886
Contingency (10%)	334,589
TOTAL ESTIMATED CAPITAL COSTS	3,680,475
	say, 3,680,500

5.4.2 Basis of Estimate

The capital cost estimate is based on the following information:

- flowsheet and conceptual layout prepared by Melis Consulting Engineers Ltd.
- budget quotations on major equipment
- pricing data from Melis Consulting Engineers Ltd. and Kilborn Ltd.'s files and records to arrive at factors for mechanical, piping, electrical and instrumentation costs. These factors have been adjusted to reflect the degree of complexity of the proposed process and the relative cost of unit equipment.

For purposes of this study it has been assumed that:

- all construction services would be tendered by fixed price contract
- the average labour rate is $35.00 per hour based on a 40-hour week. This labour rate includes provision for direct wages, payroll burdens and benefits, small tools and consumables, site preparation and the contractor's overhead and fee
- construction labour force is available in the immedediate surrounding area
- the cumulative price factor applied to base prices for process equipment is eighteen percent,which includes:
  - provincial sales tax - 7%
  - freight allowance - 5%
  - miscellaneous minor materials allowance - 5%
- all costs are reported in first quarter 1986 Canadian dollars.

No allowances have been made in the capital estimate for:

- escalation beyond the first quarter 1986
- pre-production construction interest
- working capital
- production equipment or materials
- start-up assistance

- cost of pre-appropriation studies and development work
- owners' costs such as owners'representative
- general property facilities such as administration offices
- pumping bioleach solutions from the underground leaching areas to the treatment facility
- pumping of the concentrated solution to surface with the general minewater
- decommissioning.

5.4.3 Methodology

Equipment prices were obtained on an individual basis from budget quotations and historical data. Mechanical installation costs, including field material, were factored from the equipment costs. Piping, electrical and instrumentation costs were factored from the total installed mechanical costs. Building structure costs for each area were based on volumes and volume unit prices derived from historical data. The factors used in deriving the total installed mechanical costs have been adjusted to reflect the complexity and relative cost of unit equipment for the proposed process.

The allowance for construction indirects has been estimated as eight percent of the total direct costs. It includes:

- temporary services
- permits, insurance and bonds
- rental of mobile equipment
- on-site construction supervision facility
- pre-construction geotechnical investigations.

The engineering, procurement and construction management function has been estimated as twelve percent of the direct costs. As well as design engineering, this includes provision for project management, cost control, scheduling, contract administration and inspection services.

An allowance of five percent of home office costs has been included for capital spares inventory which would include the cost of initial fills in the circuit but does not include membrane replacement.

The contingency allowance provides for additions and unidentified items within the stated scope of work but it is not intended to cover major changes in scope.

5.5 OPERATING COST ESTIMATE

5.5.1 Summary

The reverse osmosis process operating costs, based on the design basis detailed in Section 5.3, total $539,640 annually in first quarter 1986 Canadian dollars, including a five percent contingency allowance. This annual operating cost is equivalent to $\$0.92/m^3$ of bioleach solution processed or \$6.18/kg U_3O_8 (\$2.80/lb U_3O_8) based on a total of 87,325 kg U_3O_8 (192,515 lbs U_3O_8) of recovered uranium per year for the Reverse Osmosis Option.

A summary of the operating costs for a 350 day per year operation is presented in Table 5-2 with cost details given in Tables 5-3 and 5-4. The operating costs include all costs associated with the recovery of uranium from underground bioleach solutions by the proposed reverse osmosis process. The processing costs extend from receipt of the bioleach solution underground to the production of concentrated solution discharge to the main mine sump. These costs include manpower, reagents and supplies, power, and the cost of analysis for circuit operation. It does not include any costs associated with the actual underground leaching nor does it include any costs associated with downstream processing of the uranium in the surface milling circuit.

5.5.2 Manpower

The annual manpower costs detailed in Table 5-3 include the manpower estimated to be required for the operation of the reverse osmosis circuit underground. The average annual total manpower is estimated at three and a half men for meeting operation and maintenance requirements. A twenty-five percent payroll burden has been assumed in addition to the indicated salary rates.

TABLE 5-2

REVERSE OSMOSIS OPTION
ANNUAL OPERATING COSTS

Item	Annual $	$/m^3	$/kg U_3O_8	$/lb U_3O_8
Manpower	148,800	0.25	1.70	0.77
Reagents and Supplies	294,690	0.50	3.38	1.53
Power	60,450	0.10	0.69	0.32
Analysis	10,000	0.02	0.12	0.05
Sub-Total	513,940	0.87	5.89	2.67
Contingency (5%)	25,700	0.05	0.29	0.13
TOTAL	539,640	0.92	6.18	2.80

TABLE 5-3

REVERSE OSMOSIS OPTION
ANNUAL MANPOWER COSTS

Employee	Requirements	Number	Rate	Annual$
Supervision-Operation	0.5 man/day	0.5	$3,500/month	21,000
-Maintenance	0.15 man/day	0.15	$3,500/month	6,300
Operator	0.5 man/shift	2.0	$16.00/hour	67,200
Maintenance	15 mandays/month	0.75	$17.00/hour	24,500
Sub-Total		3.4		119,000
Payroll Burden (25%)				29,800
TOTAL ESTIMATED MANPOWER COSTS				148,800

TABLE 5-4

REVERSE OSMOSIS OPTION
ANNUAL REAGENTS AND SUPPLIES COSTS

Item	Unit Cost($/kg)	Consumption kg/kg U_3O_8(1)	Consumption kg/year	Annual $
Anti-scaling Agent	10.00	0.07	5,880	58,800
Sodium Hydrosulphite	2.00	0.17	15,050	30,100
Ammonium Bifluoride	2.71	0.17	15,050	40,790
Membrane Replacement(2)				50,000
Filter Bags(3)				5,000
Maintenance Supplies				75,000
Operating Supplies				30,000
Laboratory Supplies				5,000
TOTAL				294,690

Notes:

(1) Reagent consumptions are based on 87,325 kg of recovered uranium (U_3O_8) per year

(2) Membrane replacement is average annual cost based on replacement every three years.

(3) Bag replacement for bag filter assumed every two months.

The supervision allowance for operations is based on a supervisor spending fifty percent of his time on dayshift monitoring the operation of the biomass adsorption circuit. Maintenance supervision would be less with a supervisor spending approximately fifteen percent of his time in the area.

Operating manpower for the underground biomass adsorption circuit is based on an operator spending fifty percent of his time monitoring the circuit, three shifts per day, 350 days per year. A maintenance allowance of 15 mandays per month has been assumed for maintenance of the equipment in the reverse osmosis circuit.

5.5.3 Reagents and Supplies

The cost of reagents and supplies detailed in Table 5-4 is based on consumption rates given in the reverse osmosis design basis and the delivered price of reagents to the facility. The annual consumption is based on 87,325 kg of recovered uranium (U_3O_8)per year.

The anti-scaling agent comsumption of 5,880 kg per year is based on a 10 mg/L dosage rate to the reverse osmosis feed. Sodium hydrosulphite and ammonium bifluoride consumptions are based on membrane cleaning every two weeks. The $5,000 lump sum estimate for replacement of filter bags is based on the anticipated suspended solids loading at the bag filter with bag replacement being required every two months. The membrane replacement cost of $50,000 annually reflects a membrane cost of $150,000 with replacement every three years.

Lump sum estimates have been included for laboratory, maintenance and operating supplies sufficient to meet the requirement for operation of the reverse osmosis circuit.

5.5.4. Power

The total power demand for operating the reverse osmosis circuit is estimated at approximately 306 kW. The total annual power cost is thus estimated at $60,450 assuming an 85% power efficiency factor and a power cost of $0.02/kWh for a 350 day per year operation.

5.5.5 Analysis

The lump sum estimate of $10,000 for analytical costs, 1,000 analyses per year, would meet the requirements for daily metallurgical control of the reverse osmosis circuit.

5.5 TECHNICAL EVALUATION

5.6.1 Discussion

Reverse Osmosis is a proven process for upgrading brackish water for potable use. It has also been suggested as a means of recovering uranium from solutions (48,49) and for the removal of uranium in wastewater treatment (50,51,52).

The information used for this study is based for the most part on a recent limited bench-scale study (39) which provided sufficient data for detailing a proposed flowsheet for uranium recovery from bioleach solutions by reverse osmosis. The data in this report (39) was encouraging and useful for preliminary scale-up calculations but further work will be required to better define the reverse osmosis process for this application.

A reverse osmosis system would be expected to yield high uranium recoveries (99%) but a permeate/feed ratio of 0.70 would result in a solution concentration ratio of only 3.3:1. A permeate recovery of 70% is thought to be a realistic value in terms of the limited solubility of gypsum and silica in the water being processed (53).

The expected concentration of suspended solids in the bioleach solution would necessitate the use of sand filters followed by a bag filter to achieve the high clarities required by a reverse osmosis unit (53,54). The sand filter criterion is based on actual design values (45) with bag filter criterion being estimated for the anticipated sand filter discharge quality (54). A 5-micron separation for bag filtration has been assumed as satisfactory for operation of the reverse osmosis unit (54). Due to potential scaling problems an anti-scaling agent has been assumed necessary to extend membrane life (53). Consumption rates are based on a 10 mg/L dosage to the reverse osmosis feed.

The type of membrane which is felt to be most suitable for the application is Film Tec FT 30, a thin-film membrane material(39). The solution flux rate that has been assumed for this material is 0.6 $m^3/d/m^2$ (39). Cellulose acetate membranes would not be suitable due to their susceptibility to acidic solution(53). A spiral wound membrane has been assumed as the membrane configuration rather than a tubular type since there is likely a larger choice of spiral wound membranes which will operate at the low design pH(53).

An operating pressure of 2,000 kPa has been assumed in the design with an upside value of 2,750 kPa(53). Others claim that pressures as high as 3450 kPa could become necessary(39) but the lower design figure of 2,750 kPa was felt reasonable for the purpose of this study.

Membrane cleaning has been assumed to be necessary for removal of iron and silica scaling (53). For the purpose of this study frequent cleaning (bi-weekly) has been included to prolong membrane life. Actual cleaning conditions would need to be determined experimentally. There is no basis to arrive at a typical membrane life under these operating conditions so it has been assumed that a membrane would retain its efficiency for 3 years before needing replacement.

5.6.2 Recommendations

The relatively high cost of reverse osmosis units makes this option unattractive for uranium recovery from bioleach solutions. Although operating conditions could possibly be improved the cost of reverse osmosis units and the associated feed pumps, whether as small individual units or as larger units, would not change so this option would always remain capital intensive.

The lack of scale-up information for the reverse osmosis option would also necessitate a substantial amount of experimental work to provide reliable data for design purposes. The following are some suggested areas where research could be undertaken:

- the efficiency of the pre-treatment clarification step needs to be evaluated with respect to meeting manufacturer's specifications for removal of suspended solids down to 5 microns or less

- the lower limit for suspended solids removal needs to be assessed for potentially suitable membranes
- additional testing is required to determine the concentration and recovery that can be achieved under steady-state conditions
- the exact nature of the suspended solids and organic content in the bioleach solution needs to be determined in terms of membrane fouling
- the problem of gypsum, iron, and silica fouling of membranes needs to be determined as well as the effect of individual ion concentrations in the bioleach solutions on concentration efficiencies
- membrane cleaning requirements need to be quantified to determine the type of cleaning agents, cleaning conditions and cleaning frequencies as well as the beneficial effect of cleaning on membrane efficiencies
- the life expectancy of membranes when processing pH 2 solutions for extended periods is an unknown which needs to be quantified
- the effect of membrane cleaning agents (such as $Na_2S_2O_4.2H_2O$ and NH_4HF_2) on underground uranium leaching efficiencies needs to be assessed
- more studies are required to determine the type of membrane suitable for this application as well as the most suitable membrane configuration
- the extent of nutrient removal by reverse osmosis needs to be determined
- the effectiveness of an anti-scaling agent needs to be investigated.

In terms of the experimental work that seems required for confirming the suggested reverse osmosis process flowsheet a small pilot scale study would be useful. This study would only need to be on a small scale but would be operated on a continuous basis using actual bioleach solutions as opposed to synthetic feed stock. In conjunction with complimentary bench tests sufficient data would be generated to design a prototype plant if the results obtained justified a reverse osmosis process for recovering uranium from bioleach solutions.

6.0 CONCLUSIONS AND RECOMMENDATIONS

A technical comparison of the three options evaluated for uranium recovery from underground bioleach solutions is presented in Table 6-1 in terms of anticipated efficiencies for an ion exchange, biomass adsorption and reverse osmosis circuit. For the same solution throughput capacity the net annual uranium recovery is highest for the reverse osmosis option but this higher capacity is reflected in a much lower concentration ratio which in turn leads to a lower reduction in the volume of total minewater to be pumped to surface in an underground leaching program.

The highest concentration ratio, in terms of concentrated eluate production, would be obtained from an ion exchange circuit but reduction in minewater pumping is limited by the need for a barren bleed in an ion exchange circuit to control sulphate levels in the recycled barren bioleach solution.

Although circuit efficiencies can provide some insight as to which option would be most viable for in-situ uranium recovery from underground bioleach solution a much better evaluation of the three options is realized by comparing the economics of the ion exchange, biomass adsorption and reverse osmosis circuits. This economic comparison is detailed in Table 6-2.

From a capital cost point-of-view the Ion Exchange Option is by far the most attractive being less than half the cost of either the Biomass Adsorption Option or the Reverse Osmosis Option. Unit capital costs relative to annual uranium recovery and annual circuit capacity are given for each option to exemplify this cost difference. These unit capital costs are presented in terms of dollars per annual cubic meter of bioleach solution processed ($/annual m^3), dollars per annual kilogram U_3O_8 recovered ($/annual kg U_3O_8) and dollars per annual pound U_3O_8 recovered ($/annual lb U_3O_8). A capital cost ratio is also pre-

TABLE 6-1

URANIUM RECOVERY FROM BIOLEACH SOLUTIONS
TECHNICAL COMPARISON

ITEM	ION EXCHANGE	BIOMASS ADSORPTION	REVERSE OSMOSIS
Solution Flow(m^3/h)	70	70	70
Solution Headgrade(g/L U_3O_8)	0.15	0.15	0.15
Net Uranium Recovery(%)	95	81.7	99
Uranium Recovered(kg U_3O_8/year)	83,790	72,100	87,325
Concentration Ratio	32:1	4.7:1	3.3:1
% Minewater Pumping Reduction	84	86	70

TABLE 6-2

URANIUM RECOVERY FROM BIOLEACH SOLUTIONS
ECONOMIC COMPARISON
(FIRST QUARTER 1986 CANADIAN DOLLARS)

	ION EXCHANGE	BIOMASS ADSORPTION	REVERSE OSMOSIS
Capital Costs			
Total $	1,674,400	3,736,600	3,680,500
$/annual m^3 processed	2.85	6.35	6.26
$/annual kg U_3O_8 recovered	19.98	51.83	42.15
$/annual lb U_3O_8 recovered	9.06	23.51	19.12
Relative Cost- m^3 processed	1.0	2.2	2.2
- kg U_3O_8 recovered	1.0	2.6	2.1
Operating Costs			
Annual $	374,300	3,519,820	539,640
$/$m^3$ processed	0.64	6.00	0.92
$/kg U_3O_8 recovered	4.46	48.83	6.18
$/lb U_3O_8 recovered	2.03	22.13	2.80
Relative Cost- m^3 processed	1.0	9.4	1.4
-kg U_3O_8 recovered	1.0	10.9	1.4
Total Costs*			
Total $	5,417,400	38,934,800	9,076,900
$/$m^3$ processed	0.92	6.62	1.54
$/kg U_3O_8 recovered	6.47	54.00	10.39
$/lb U_3O_8 recovered	2.93	24.49	4.71
Relative Cost- m^3 processed	1.0	7.2	1.7
- kg U_3O_8 recovered	1.0	8.3	1.6

* based on an assumed 10-year operating life

sented for the Biomass Adsorption Option and Reverse Osmosis Option relative to the Ion Exchange Option both in terms of throughput capacity and uranium recovery.

The above capital cost assessment highlights the significantly lower capital cost of the Ion Exchange Option over the Biomass Adsorption and Reverse Osmosis Options. It should be noted that these capital cost estimates do not include any provision for pre-production construction interest since the cost of this interest is dependent on the method of project financing and the equity or credit position of the specific investor. These pre-production interest costs can have a significant impact on the total costs of any project and this should be kept in mind when reviewing the cost estimates given in this report.

The annual operating costs of each option are summarized in Table 6-2. These costs include all the associated costs of operating the uranium recovery circuit for each option with a bioleach solution throughput capacity of 70 m^3/h or 588,000 m^3 per year. It should be noted that the operating cost of the Ion Exchange Option includes the production of concentrated eluate (yellowcake precipitation feed) on surface whereas the concentrated solutions for the other two options would still need to be processed through the main ion exchange circuit in the mill facility to provide feed liquor for yellowcake precipitation. The additional operating cost for this would be minimal however relative to the estimated operating cost of the Biomass Adsorption Option and Reverse Osmosis Option for producing concentrated solution underground.

In Table 6-2 unit operating costs are given in terms of dollars per cubic meter ($\$/m^3$) of bioleach solution processed, dollars per kilogram U_3O_8($\$/kg\ U_3O_8$) recovered, and dollars per pound U_3O_8($\$/lb\ U_3O_8$) recovered. An operating cost ratio is also presented for the Biomass Adsorption Option and Reverse Osmosis Option relative to the Ion Exchange Option both in terms of throughput capacity and uranium recovery. This operating cost comparison shows that the Biomass

Adsorption Option would be approximately ten times higher in operating costs than ion exchange while the Reverse Osmosis Option would be approximately one and a half times more costly to operate.

An indication of the cost difference between the three options when combining both the capital costs (without amortization) and operating costs can be obtained by calculating the total costs of each option for an assumed 10-year operating life at the 70 m^3/h rated capacity using the expected uranium recovery for each option. This cost comparison is given in Table 6-2 in terms of total costs for each option and as unit costs expressed as dollars per cubic meter ($\$/m^3$) of bioleach solution processed, dollars per kilogram U_3O_8 ($\$/kg\ U_3O_8$) recovered and dollars per pound U_3O_8 ($\$/lb\ U_3O_8$) recovered. The total volume of bioleach solution processed in a 10-year operation would be the same for each option, namely 5,880,000 m^3, while the total uranium recovery for the 10-year operation would be related to the uranium recovery efficiency of each option. A total cost ratio is also presented for the Biomass Adsorption Option and Reverse Osmosis Option relative to the Ion Exchange Option both in terms of operating throughout capacity and uranium recovery.

On a total cost basis the Biomass Adsorption Option would be seven times more expensive than the Ion Exchange Option in terms of bioleach solution volume processed and eight times more costly in terms of uranium recovered. The total costs estimated for the Reverse Osmosis Option is approximately one and a half times more expensive than the Ion Exchange Option both in terms of solution processed and uranium recovered.

In summary then the lower capital and operating cost of the Ion Exchange Option, in conjunction with its acceptibility from a practical standpoint as an established process for uranium recovery, based on

current technology, makes this option the most suitable process for in-situ uranium recovery from underground bioleach solutions.

The high costs of the Biomass Adsorption Option are attributed to the extent of process equipment required for the conceptual biomass adsorption circuit derived from available current technology for this option. The relatively low uranium loading efficiency of biomass along with its limited recycling potential and inherent high production cost also contribute significantly to the total costs of this option. Pertinent research areas have been identified to reduce the costs of this option and to improve the viability of using biomass adsorption in hydrometallurgical processes such as uranium recovery from underground bioleach solutions.

The Reverse Osmosis Option appears to be more favourable in terms of total costs. However, the cost of reverse osmosis units and the associated high-pressure feed pumps are the major reason for the high cost of this option relative to the Ion Exchange Option and hence there appears to be little potential for making this option more economically attractive. Solubility concerns with respect to the chemical make-up of bioleach solution would also limit the extent of uranium concentration in upgraded bioleach solutions which could be obtained in a reverse osmosis circuit. Hence the potential for significantly reducing minewater pumping requirements in an underground bioleach operation would always be less than what could be possible from an ion exchange or biomass adsorption circuit.

In conclusion the technical and economic evaluation detailed in this report has shown ion exchange to be the most favourable processing option of the three options evaluated for in-situ uranium recovery from underground bioleach solutions. Since the current method of recovering uranium from minewater in uranium operations is by precipitation with lime and subsequent dissolution of the recovered precipitate the potential for using lime precipitation underground as a means of

recovering uranium from bioleach solution should be investigated. This precipitation alternative would include pumping the resulting precipitate to surface as a dilute slurry for final uranium recovery in the mill circuit. The viability of this option could then be assessed and compared against the Ion Exchange Option to arrive at the most suitable method for in-situ uranium recovery from underground bioleach solutions.

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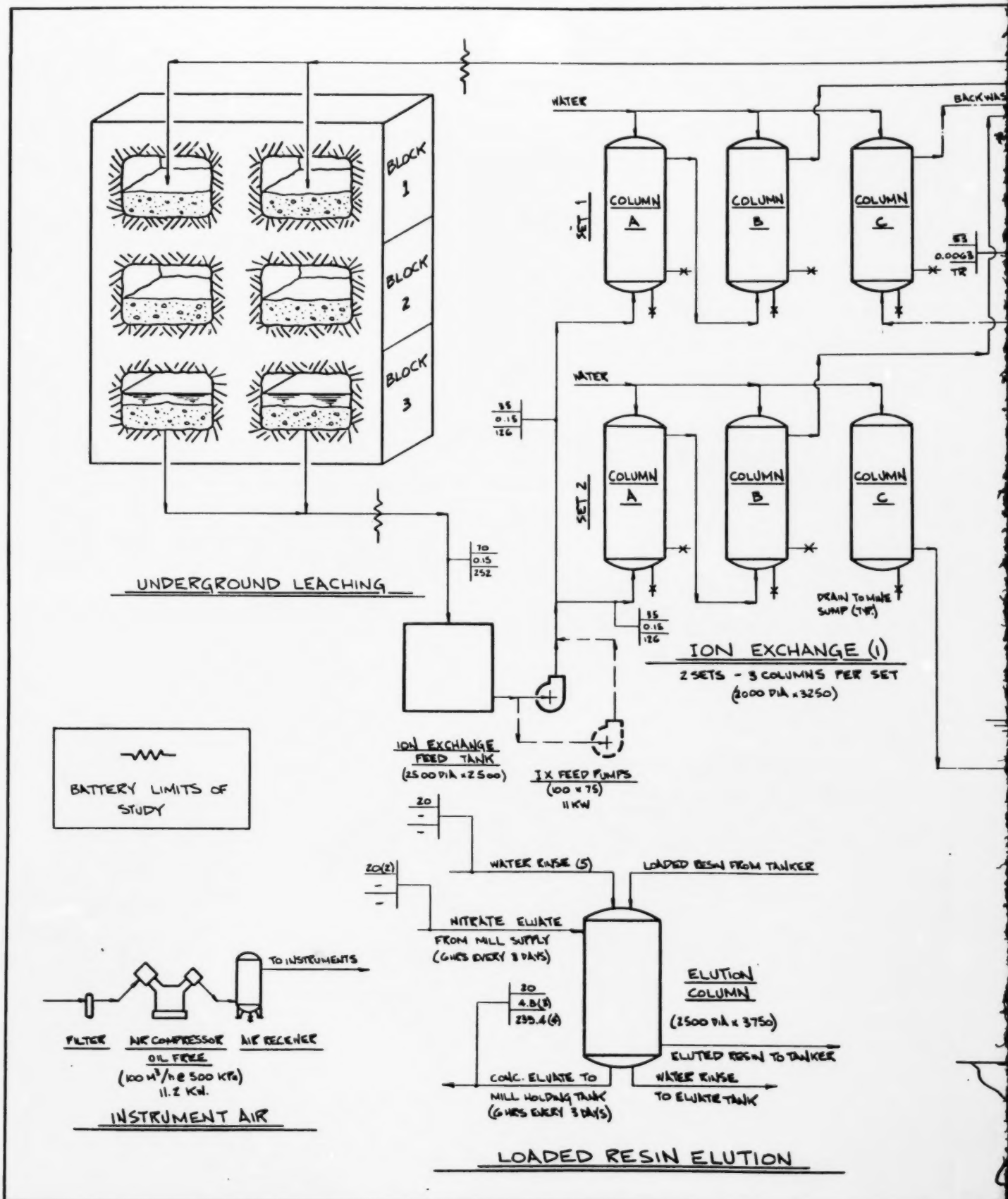
8.0 DRAWINGS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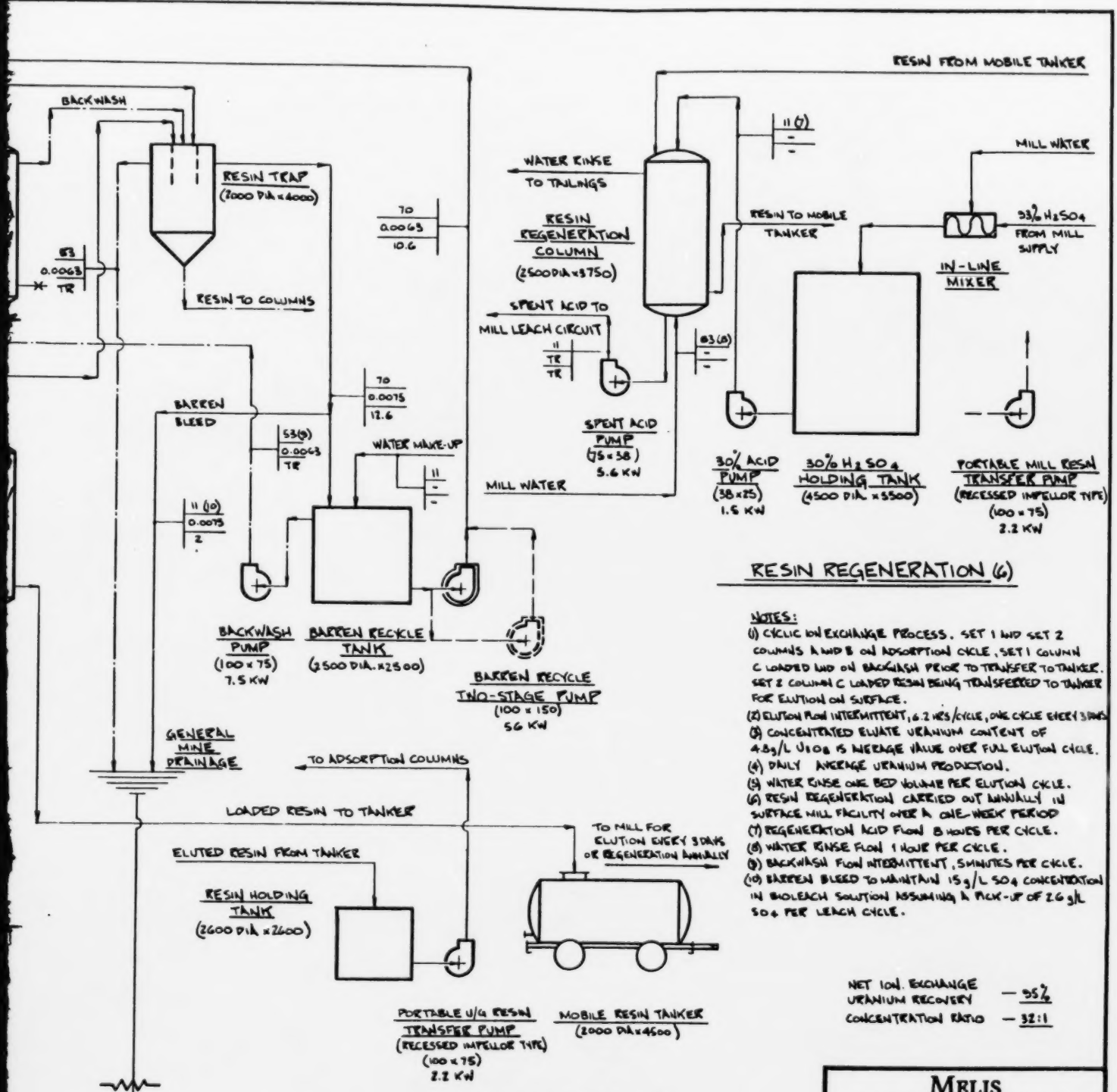
BLOCK 2
BLOCK 3
UNDERGROUND LEACHING
WATER
SET 1
COLUMN A
COLUMN B
COLUMN C
BACKWASH
E3
0.0063
TR
WATER
SET 2
COLUMN A
COLUMN B
COLUMN C
DRAIN TO MAKE SUMP (TYP.)
35
0.15
126
70
0.15
252
ION EXCHANGE (1)
2 SETS - 3 COLUMNS PER SET
(2000 DIA x 3250)
ION EXCHANGE FEED TANK
(2500 DIA x 2500)
IX FEED PUMPS
(100 x 75)
11 KW
BATTERY LIMITS OF STUDY
20
20(2)
WATER RINSE (5)
LOADED RESIN FROM TANKER
NITRATE ELUATE
FROM MILL SUPPLY
(6 HRS EVERY 3 DAYS)
ELUTION COLUMN
(2500 DIA x 3750)
20
4.8(8)
239.4(6)
ELUTED RESIN TO TANKER
CONC. ELUATE TO MILL HOLDING TANK
(6 HRS EVERY 3 DAYS)
WATER RINSE TO ELUATE TANK
LOADED RESIN ELUTION
TO INSTRUMENTS
FILTER
AIR COMPRESSOR
OIL FREE
(100 m³/hr @ 500 KPa)
11.2 KW.
AIR RECEIVER
INSTRUMENT AIR



RESIN REGENERATION (6)

NOTES:

(1) CYCLIC ION EXCHANGE PROCESS. SET 1 AND SET 2 COLUMNS A AND B ON ADSORPTION CYCLE, SET 1 COLUMN C LOADED AND ON BACKWASH PRIOR TO TRANSFER TO TANKER. SET 2 COLUMN C LOADED RESIN BEING TRANSFERRED TO TANKER FOR ELUTION ON SURFACE.

(2) ELUTION FLOW INTERMITTENT, 6.2 HRS/CYCLE, ONE CYCLE EVERY 3 DAYS

(3) CONCENTRATED ELUATE URANIUM CONTENT OF 4.8 g/L U_3O_8 IS AVERAGE VALUE OVER FULL ELUTION CYCLE.

(4) DAILY AVERAGE URANIUM PRODUCTION.

(5) WATER RINSE ONE BED VOLUME PER ELUTION CYCLE.

(6) RESIN REGENERATION CARRIED OUT ANNUALLY IN SURFACE MILL FACILITY OVER A ONE-WEEK PERIOD

(7) REGENERATION ACID FLOW 8 HOURS PER CYCLE.

(8) WATER RINSE FLOW 1 HOUR PER CYCLE.

(9) BACKWASH FLOW INTERMITTENT, 5 MINUTES PER CYCLE.

(10) BARREN BLEED TO MAINTAIN 15 g/L SO_4 CONCENTRATION IN BIOLEACH SOLUTION ASSUMING A PICK-UP OF 2.6 g/L SO_4 PER LEACH CYCLE.

NET ION EXCHANGE URANIUM RECOVERY — 95%

CONCENTRATION RATIO — 32:1

RESIN TRANSFER

PERIODIC FLOW

ALTERNATE FLOW

m^3/h

U_3O_8 g/L

U_3O_8 Kg/DAY

NO.	DATE	REVISION	BY	APP.
1	26 SEP 86	TANK SIZES REDUCED & CHANGED TO INTERMITTENT ELUTION OF RESIN ON SURFACE	N.S.	LAM
2	6 NOV 86	CHANGE NAME OF ONE TANK TO ION EXCHANGE FEED TANK	N.S.	

SCALE N.T.S. DATE

REVISED LAM Dec. '85

DRAWN Shad Dec. '85

CHECKED

APPROVED

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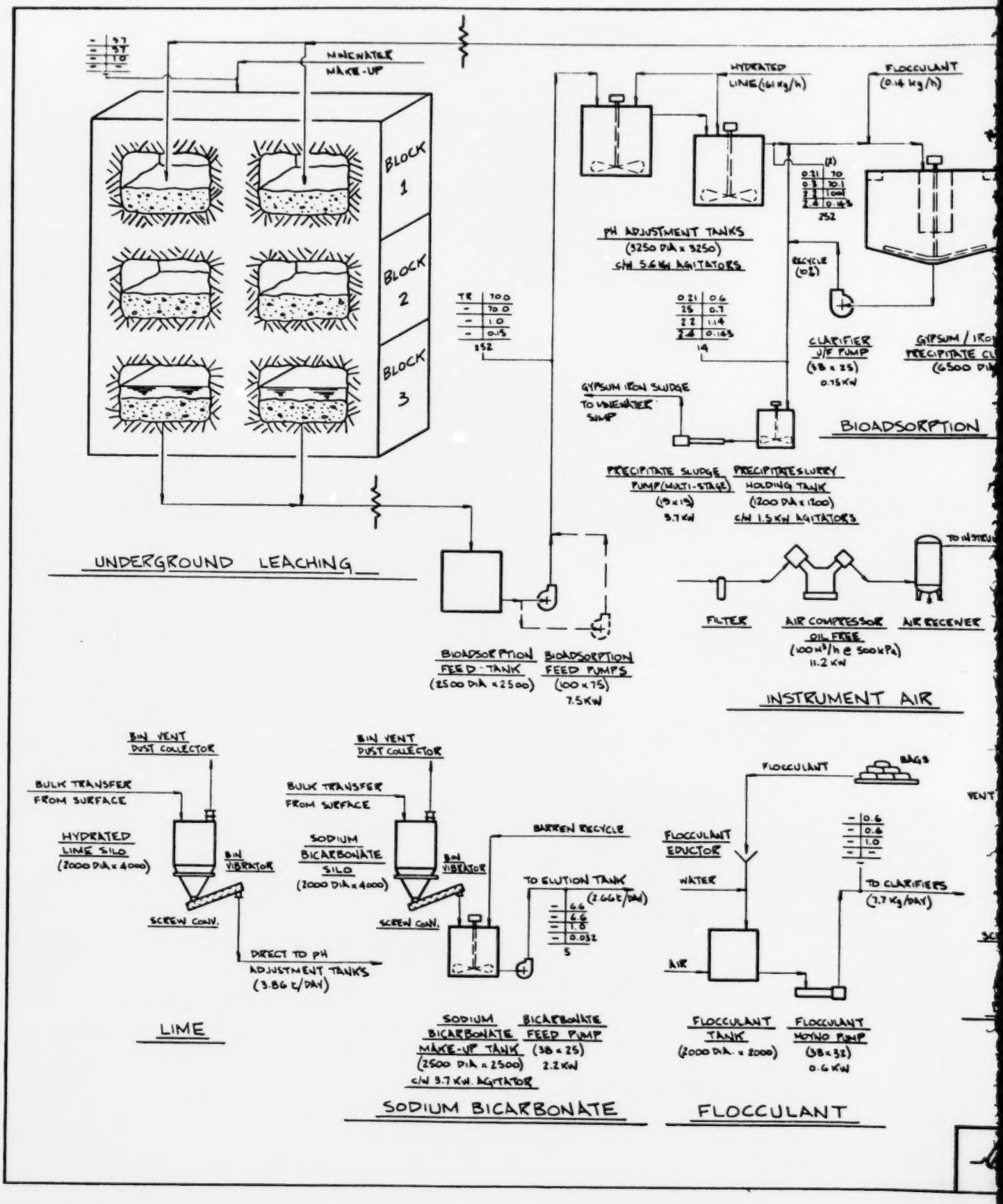
CLIENT Dept. Energy, Mines & Resources
Ottawa Ontario

URANIUM BIOLEACH STUDY

PROCESS FLOWSHEET

ION EXCHANGE OPTION

PROJECT NO. MCEL - 108 | DRAWING NO. 100-10-1 | REV. 2


MINEWATER MAKE-UP
BLOCK 1
BLOCK 2
BLOCK 3
UNDERGROUND LEACHING
HYDRATED LIME (61 kg/h)
FLOCCULANT (0.14 kg/h)
PH ADJUSTMENT TANKS (3250 DIA x 3250) c/w 5.6 KW AGITATORS
RECYCLE (0%)
CLARIFIER U/F PUMP (38 x 25) 0.75 KW
GYPSUM IRON SLUDGE TO MINEWATER SUMP
BIOADSORPTION
PRECIPITATE SLUDGE PUMP (MULTI-STAGE) (19 x 13) 3.7 KW
PRECIPITATE SLURRY HOLDING TANK (1200 DIA x 1200) c/w 1.5 KW AGITATORS
BIOADSORPTION FEED TANK (2500 DIA x 2500)
BIOADSORPTION FEED PUMPS (100 x 75) 7.5 KW
FILTER
AIR COMPRESSOR OIL FREE (100 m³/h @ 500 kPa) 11.2 KW
AIR RECEIVER
INSTRUMENT AIR
BIN VENT DUST COLLECTOR
BULK TRANSFER FROM SURFACE
HYDRATED LIME SILO (2000 DIA x 4000)
BIN VIBRATOR
SCREW CONV.
DIRECT TO PH ADJUSTMENT TANKS (3.86 L/DAY)
LIME
SODIUM BICARBONATE SILO (2000 DIA x 4000)
BARREN RECYCLE
TO ELUTION TANK (2.66 L/DAY)
SODIUM BICARBONATE MAKE-UP TANK (2500 DIA x 2500) c/w 3.7 KW AGITATOR
BICARBONATE FEED PUMP (38 x 25) 2.2 KW
SODIUM BICARBONATE
FLOCCULANT
BAGS
FLOCCULANT EDUCTOR
WATER
AIR
TO CLARIFIERS (7.7 Kg/DAY)
FLOCCULANT TANK (2000 DIA. x 2000)
FLOCCULANT MOYNO PUMP (38 x 32) 0.6 KW

FLOCCULANT (0.18 kg/h)
(3) BIOMASS MAKE-UP (0.05 t/h)
BIOADSORPTION TANK (4) (4000 DIA x 4000) c/w 5.6KW AGITATOR
GYPSUM / IRON PRECIPITATE CLARIFIER (6500 DIA.)
RECYCLE (10%)
TO SODIUM BICARBONATE MAKE-UP TANK (6)
BIOMASS CLARIFIER U/F PUMP (75 x 50) 3.7 KW
BIOMASS CLARIFIER (10000 DIA.)
BARREN RECYCLE TANK (2500 DIA x 2500)
BARREN RECYCLE PUMPS (TWO-STAGE) (100 x 150) 56 KW
TO INSTRUMENTS
BLEED TO MINE SUPPLY
LOADED BIOMASS CENTRIFUGE (610 x 2490) 75 KW
SODIUM BICARBONATE SOLUTION
VENT
DESORBED BIOMASS CENTRIFUGE (432 x 1270) 37 KW
ELUTION TANK (1600 DIA x 1600) c/w 3.7 kW AGITATOR
ELUTION PUMP (75 x 38) 3.7 KW
GYPSUM / IRON SLUDGE
BIOMASS (ENTOSPUS ARRHIZUS) FROM OFF-SITE SUPPLIER
BIOMASS STORAGE BIN (2500 DIA x 4000)
SCREW CONV.
DIRECT TO BIOADSORPTION TANK (1.2 t/DAY)
BY GRAVITY TO MINE SUMP
ELUATE SUMP (1200 DIA x 1200)
TO SURFACE FOR URANIUM RECOVERY IN MILL CIRCUIT
ELUTION
BIOMASS (7)
MAIN MINE SUMP

NOTES:

(1) LIME CONSUMPTION ESTIMATED FOR TARGET PH OF 8.5.

(2) PRECIPITATE PRODUCTION ESTIMATED BASED ON DATA FOR SIMILAR SOLUTIONS.

(3) BIOMASS MAKE-UP REQUIREMENT BASED ON 10 ADSORPTION/ELUTION CYCLES FOR THE BIOMASS.

(4) STIRRED REACTOR FOR ADSORPTION ASSUMED AS OPERATING SYSTEM BASED ON CURRENT, KNOWN TECHNOLOGY.

(5) BIOMASS S.G. ASSUMED AS 1.0.

(6) BARREN RECYCLE IS USED FOR BICARBONATE MAKE-UP TO MINIMIZE Na_2HCO_3 CONSUMPTION

(7) BIOMASS IS ASSUMED TO BE MANUFACTURED OFF-SITE AND RECEIVED IN BULK IN A DRY FORM.

NET BIOADSORPTION URANIUM RECOVERY — 81.7%
CONCENTRATION RATIO — 4.7:1

LEGEND

t/h SOLID	m³/h LIQUID
% SOLIDS	m³/h PULP
SG SOLIDS	SG PULP
kg/t U_3O_8 in SOLID	g/L U_3O_8 in LIQUID
U_3O_8 kg/DAY	

BATTERY LIMITS OF STUDY

NO.	DATE	REVISION	BY	APP.
1	Nov 86	TANK SIZES REDUCED, FLOWS ADJUSTED FOR U_3O_8 IN GYPSUM/IRON	N.S.	L.A.M.
2	Nov 86	CHANGED KW RATING ON TWO PUMPS	N.S.	

SCALE N.T.S. DATE
DESIGNED L.A.M. Dec. '85
DRAWN Shad Dec. '85
CHECKED
APPROVED

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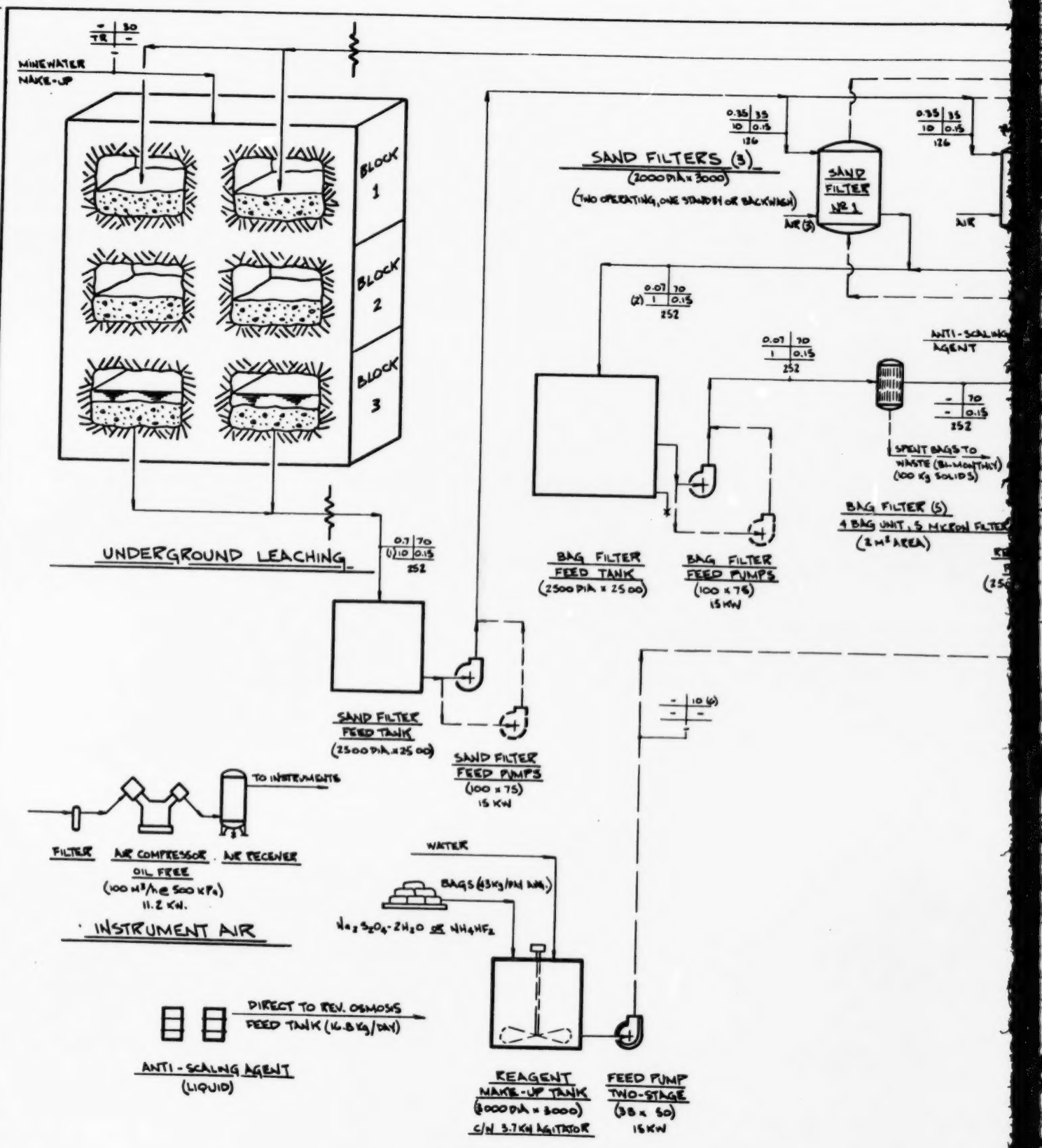
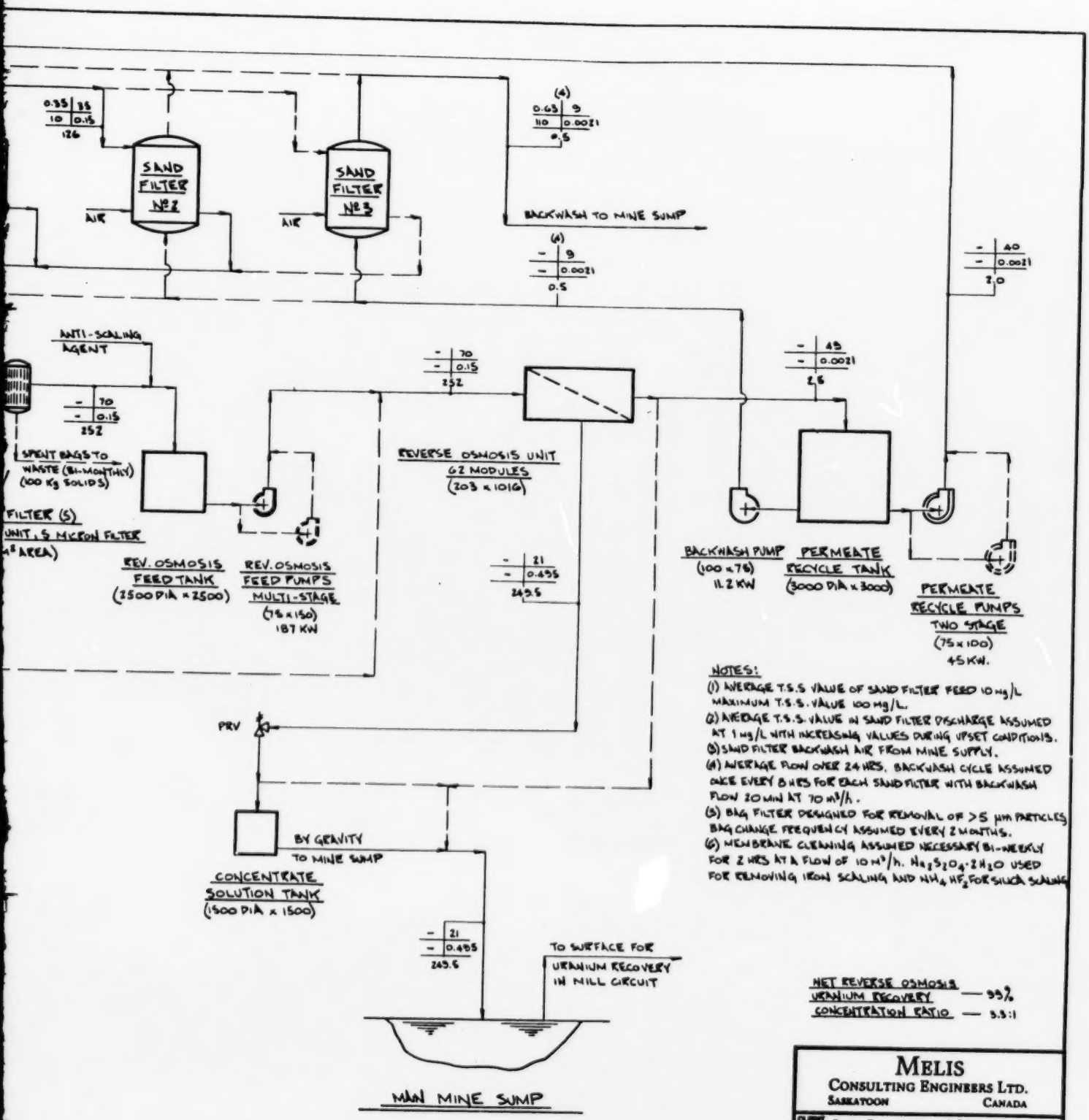
CLIENT Dept. Energy, Mines & Resources
Ottawa Ontario

URANIUM BIOLEACH STUDY
PROCESS FLOWSHEET
BIOMASS ADSORPTION OPTION

PROJECT NO. MCEL - 108 DRAWING NO. 100-10-2 REV. 2


MINEWATER MAKE-UP
BLOCK 1
BLOCK 2
BLOCK 3
UNDERGROUND LEACHING
SAND FILTERS (3)
(2000 DIA x 3000)
(TWO OPERATING, ONE STANDBY OR BACKWASH)
SAND FILTER NR 1
AIR (3)
AIR
ANTI-SCALING AGENT
SPENT BAGS TO WASTE (BI-MONTHLY) (100 Kg SOLIDS)
BAG FILTER (5)
4 BAG UNIT, 5 MICRON FILTER
(2 M² AREA)
BAG FILTER FEED TANK
(2500 DIA x 2500)
BAG FILTER FEED PUMPS
(100 x 75)
15 KW
SAND FILTER FEED TANK
(2500 DIA x 2500)
SAND FILTER FEED PUMPS
(100 x 75)
15 KW
TO INSTRUMENTS
FILTER
AIR COMPRESSOR
OIL FREE
(100 m³/hr @ 500 kPa)
11.2 KW.
AIR RECEIVER
INSTRUMENT AIR
WATER
BAGS (43 Kg/DM MIN.)
$Na_2S_2O_4 \cdot 2H_2O$ OR NH_4HF_2
DIRECT TO REV. OSMOSIS FEED TANK (16.8 Kg/DAY)
ANTI-SCALING AGENT
(LIQUID)
REAGENT MAKE-UP TANK
(2000 DIA x 3000)
C/W 3.7 KW AGITATOR
FEED PUMP TWO-STAGE
(38 x 50)
15 KW
MEMBRANE CLEANING AGENT


SAND FILTER №2
SAND FILTER №3
AIR
AIR
BACKWASH TO MINE SUMP
ANTI-SCALING AGENT
SPENT BAGS TO WASTE (BI-MONTHLY) (100 Kg SOLIDS)
FILTER (S)
UNIT, 5 MICRON FILTER
m² AREA)
REV. OSMOSIS FEED TANK (2500 DIA x 2500)
REV. OSMOSIS FEED PUMPS MULTI-STAGE (75 x 150) 187 KW
REVERSE OSMOSIS UNIT 62 MODULES (203 x 1016)
BACKWASH PUMP (100 x 75) 11.2 KW
PERMEATE RECYCLE TANK (3000 DIA x 3000)
PERMEATE RECYCLE PUMPS TWO STAGE (75 x 100) 45 KW.
PRV
BY GRAVITY TO MINE SUMP
CONCENTRATE SOLUTION TANK (1500 DIA x 1500)
TO SURFACE FOR URANIUM RECOVERY IN MILL CIRCUIT
MAIN MINE SUMP

NOTES:

(1) AVERAGE T.S.S VALUE OF SAND FILTER FEED 10 mg/L MAXIMUM T.S.S. VALUE 100 mg/L.

(2) AVERAGE T.S.S. VALUE IN SAND FILTER DISCHARGE ASSUMED AT 1 mg/L WITH INCREASING VALUES DURING UPSET CONDITIONS.

(3) SAND FILTER BACKWASH AIR FROM MINE SUPPLY.

(4) AVERAGE FLOW OVER 24 HRS. BACKWASH CYCLE ASSUMED ONCE EVERY 8 HRS FOR EACH SAND FILTER WITH BACKWASH FLOW 20 MIN AT 70 m³/h.

(5) BAG FILTER DESIGNED FOR REMOVAL OF >5 μm PARTICLES BAG CHANGE FREQUENCY ASSUMED EVERY 2 MONTHS.

(6) MEMBRANE CLEANING ASSUMED NECESSARY BI-WEEKLY FOR 2 HRS AT A FLOW OF 10 m³/h. $Na_2S_2O_4 \cdot 2H_2O$ USED FOR REMOVING IRON SCALING AND NH_4HF_2 FOR SILICA SCALING

NET REVERSE OSMOSIS URANIUM RECOVERY — 99%
CONCENTRATION RATIO — 5.5:1

BATTERY LIMITS OF STUDY

LEGEND
Kg/h SOLIDS | m³/h
TSS mg/L | g/L U_3O_8
Kg/day U_3O_8

NO.	DATE	REVISION	BY	APP.
1	86 08	TANK SIZE REDUCED ONE REAGENT TANK ELIMINATED	N.S.	LAM

SCALE	N.T.S.
DESIGNED	L.A.M. Dec. '85
DRAWN	Shad Dec. '85
CHECKED	
APPROVED	

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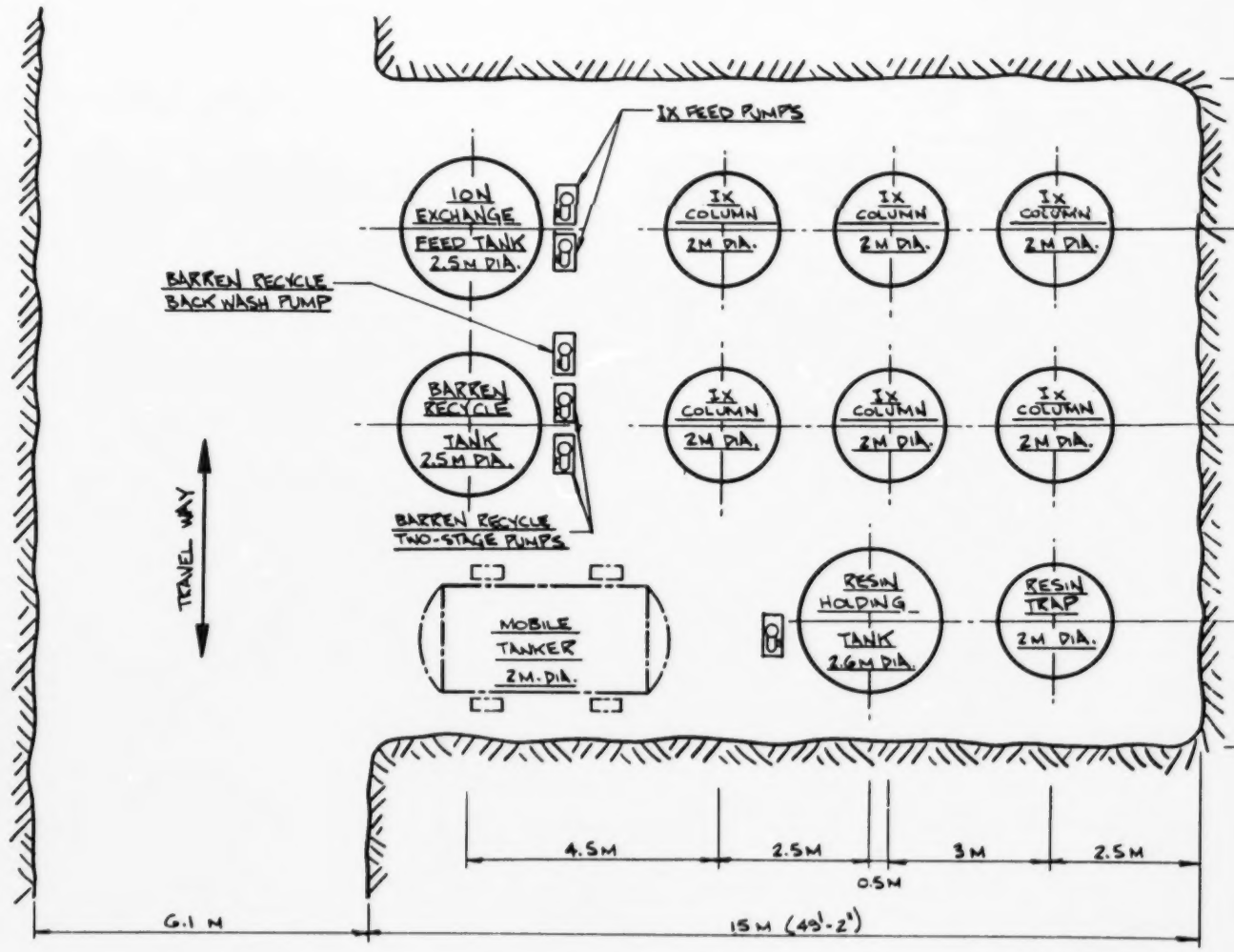
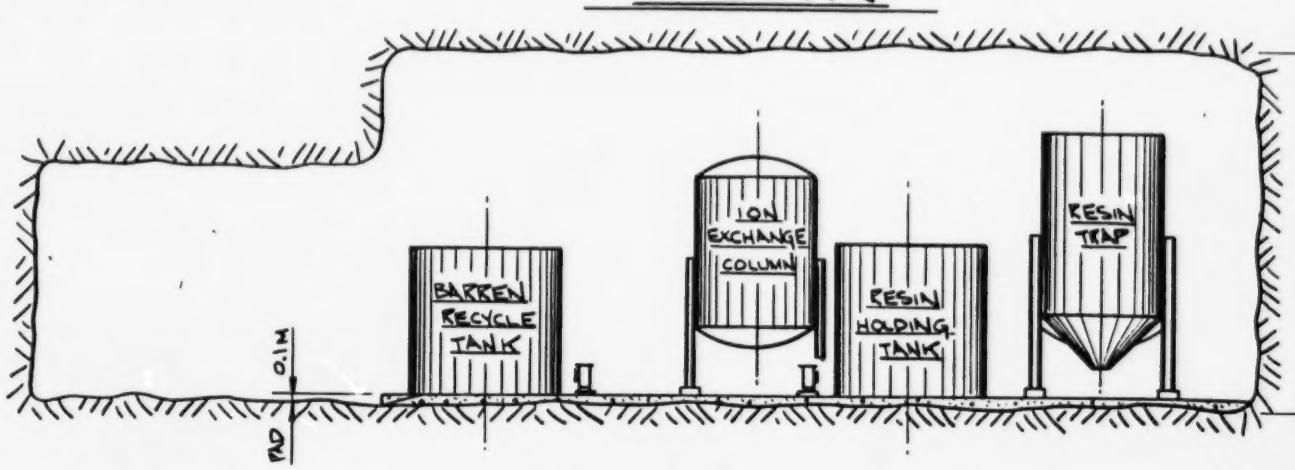
CLIENT: Dept. Energy, Mines & Resources Ottawa Ontario

URANIUM BIOLEACH STUDY
PROCESS FLOWSHEET
REVERSE OSMOSIS OPTION

PROJECT NO. MCEL - 108
DRAWING NO. 100-10-3
REV. 1



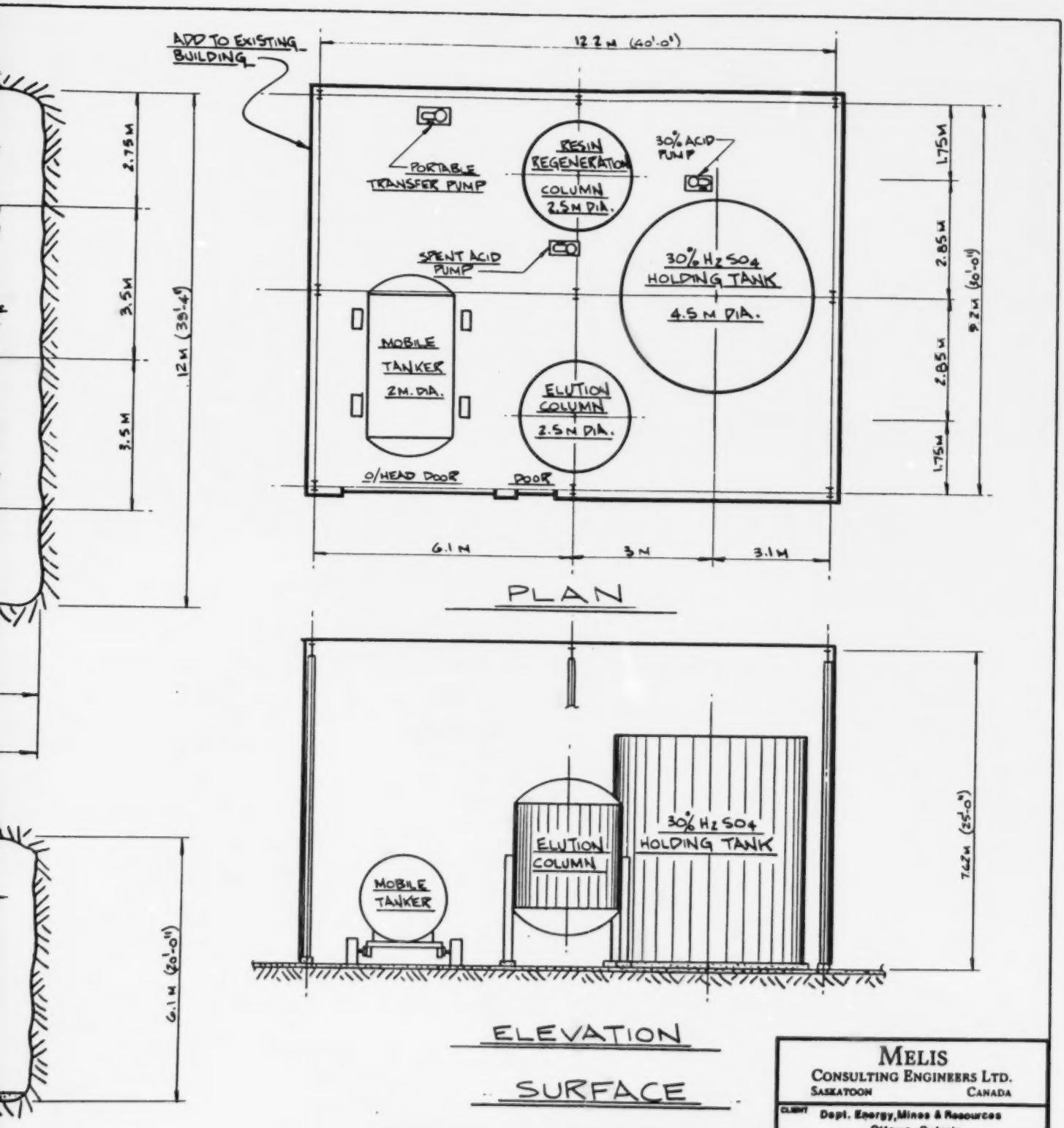
PLAN



ELEVATION

UNDERGROUND

SCALE



NO.	DATE	REVISION	BY	APP.

SCALE	M.T.S.	DATE
DESIGNED	L.A.M.	Mar. '86
DRAWN	Shad	Mar. '86
CHECKED		
APPROVED		

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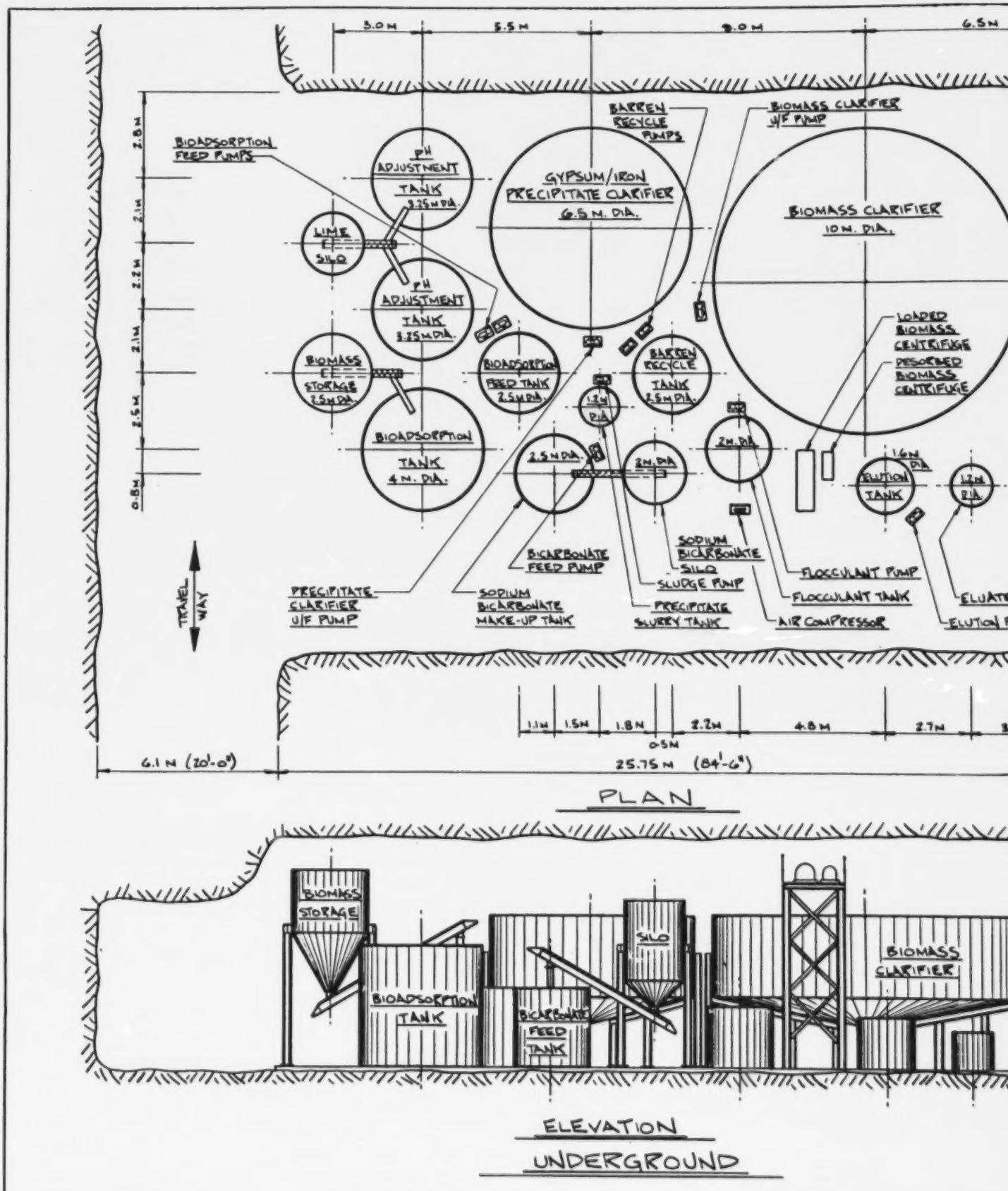
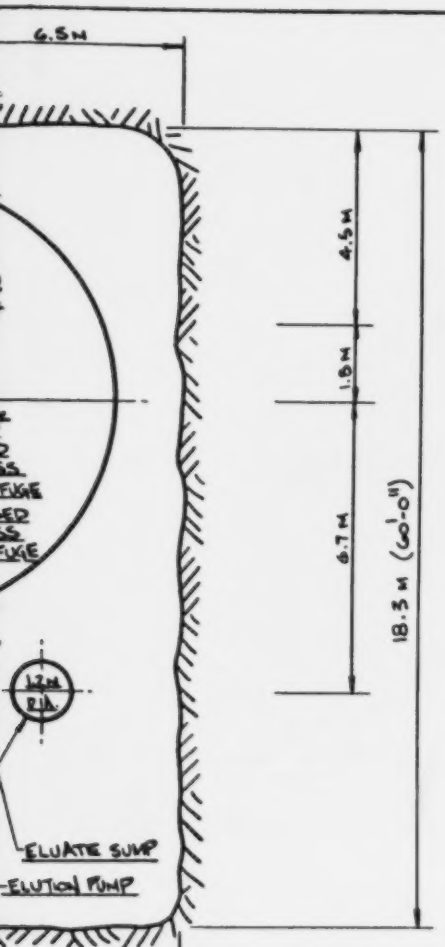
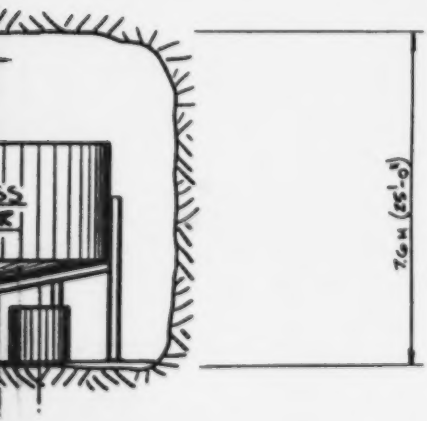
CLIENT: Dept. Energy, Mines & Resources
Ottawa, Ontario

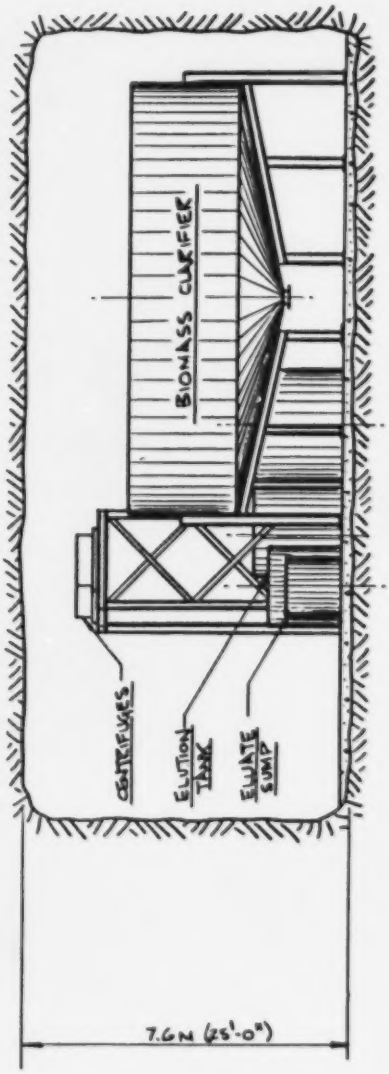
URANIUM BIOLEACH STUDY
CONCEPTUAL LAYOUT
ION EXCHANGE OPTION

PROJECT NO.	DRAWING NO.	REV.
MCEL - 108	100-20-1	0


3.0 M
5.5 M
9.0 M
6.5 M
2.8 M
2.1 M
2.2 M
2.1 M
2.5 M
0.8 M
BIOADSORPTION FEED PUMPS
PH ADJUSTMENT TANK 3.25 M DIA.
LIME SILO
PH ADJUSTMENT TANK 3.25 M DIA.
BIOMASS STORAGE 2.5 M DIA.
BIOADSORPTION TANK 4 M. DIA.
GYPSUM/IRON PRECIPITATE CLARIFIER 6.5 M. DIA.
BARREN RECYCLE PUMPS
BIOADSORPTION FEED TANK 2.5 M DIA.
BARREN RECYCLE TANK 2.5 M DIA.
1.2 M DIA.
2.5 M DIA.
2 M. DIA.
2 M. DIA.
BIOMASS CLARIFIER U/F PUMP
BIOMASS CLARIFIER 10 M. DIA.
LOADED BIOMASS CENTRIFUGE
DESORBED BIOMASS CENTRIFUGE
1.6 M DIA.
ELUTION TANK
1.2 M DIA.
TRAVEL WAY
PRECIPITATE CLARIFIER U/F PUMP
SODIUM BICARBONATE MAKE-UP TANK
BICARBONATE FEED PUMP
SODIUM BICARBONATE SILO
SLUDGE PUMP
PRECIPITATE SLURRY TANK
FLOCCULANT PUMP
FLOCCULANT TANK
AIR COMPRESSOR
ELUATE
ELUTION
1.1 M
1.5 M
1.8 M
2.2 M
0.5 M
4.8 M
2.7 M
6.1 M (20'-0")
25.75 M (84'-6")
PLAN
BIOMASS STORAGE
BIOADSORPTION TANK
BICARBONATE FEED TANK
SILO
BIOMASS CLARIFIER
ELEVATION
UNDERGROUND







ELEVATION

1 0 1 2

SCALE : METERS

NO.	DATE	REVISION	BY	APP.

SCALE	N.T.S.
DESIGNED	L.A.M.
DRAWN	Shad
CHECKED	
APPROVED	
DATE	Mar '88
DATE	Mar '88

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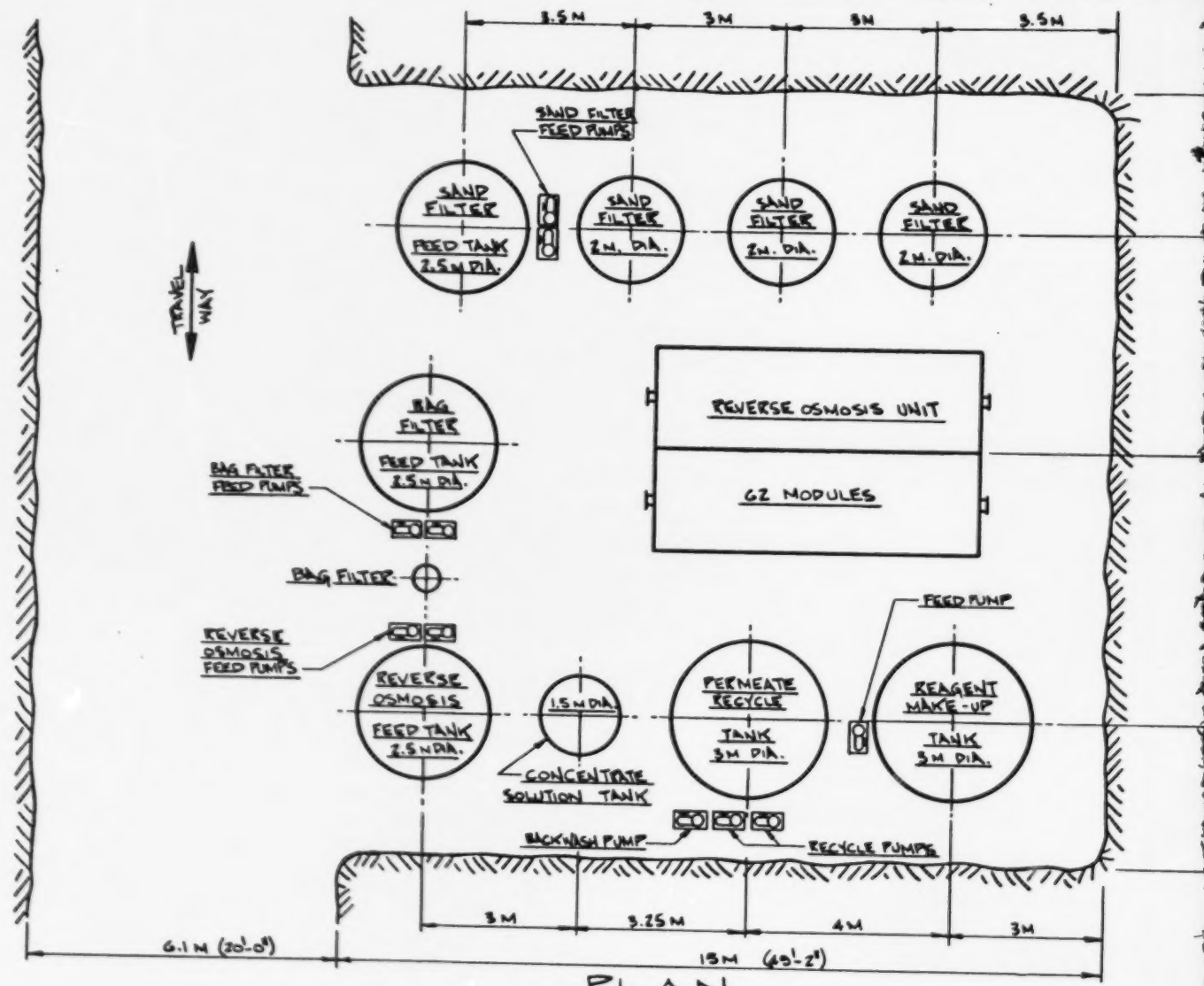
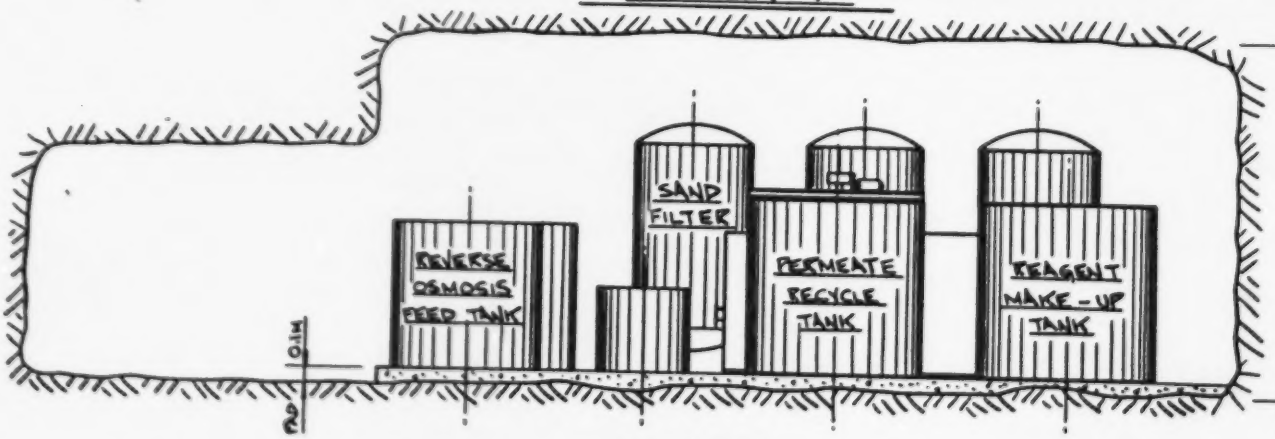
CLIENT: Dept. Energy, Mines & Resources
Ottawa, Ontario

URANIUM BIOLEACH STUDY
CONCEPTUAL LAYOUT
BIOMASS ADSORPTION OPTION

PROJECT NO.	DRAWING NO.	REV.
MCEL - 188	100-20-2	0

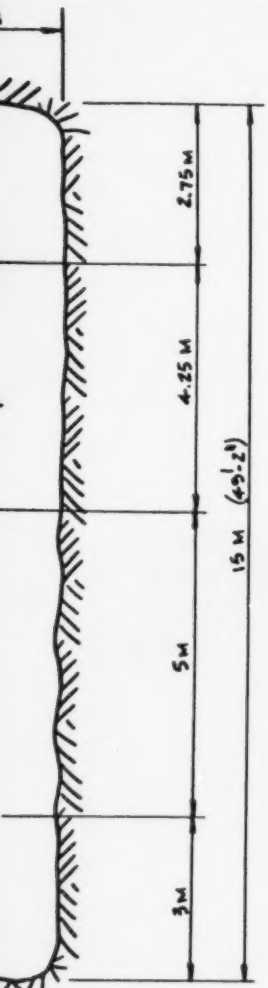


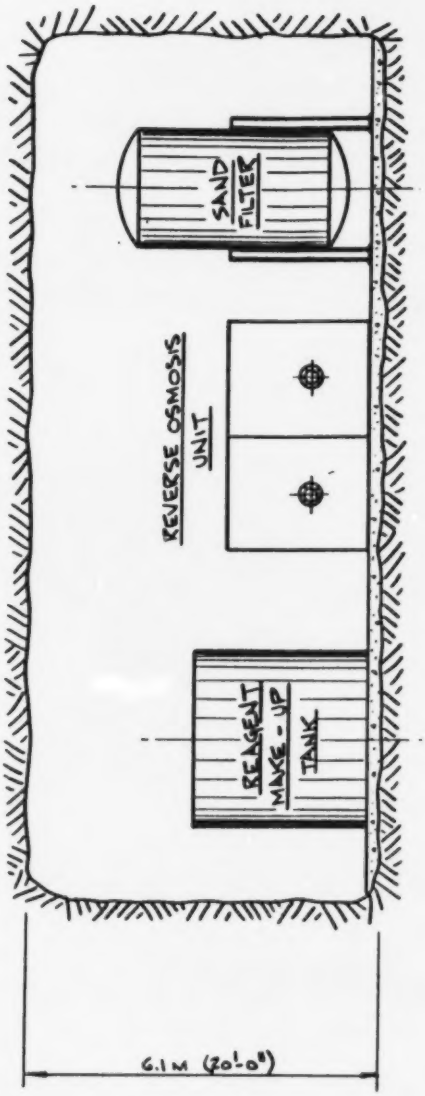
PLAN



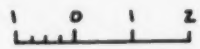
ELEVATION

UNDERGROUND





ELEVATION



SCALE : METERS

NO.	DATE	REVISION	BY	APP.

SCALE	N.T.S.
DESIGNED	L.A.M. Mar. '86
DRAWN	Shad Mar. '86
CHECKED	
APPROVED	

MELIS
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SASKATOON CANADA

CLIENT: Dept. Energy, Mines & Resources
Ottawa, Ontario

URANIUM BIOLEACH STUDY
CONCEPTUAL LAYOUT
REVERSE OSMOSIS OPTION

PROJECT NO.	DRAWING NO.	REV.
MCEL - 108	100-20-3	0

APPENDIX A

ORIGINAL STATEMENT OF WORK

APPENDIX "A"

14SQ.23440-3-9042
0SQ85-00159

STATEMENT OF WORK

REQUIREMENT

On behalf of the Department of Energy, Mines and Resources, to prepare a report which critically examines and provides technical and economic comparisons of, suitable flowsheets for the recovery of uranium from bioleach solution by the use of biomass, ion-exchange and reverse osmasis.

The flowsheets necessary for the evaluation shall be determined by the contractor based on literature already available on recovery of uranium by the use of biomass, ion-exchange and reverse osmosis respectively.

The critical evaluation shall clearly state the assumptions made, if any, and the basis and reasoning for the assumptions so made. If on the judgement of the contractor these assumptions have to be checked and verified by experiments, the contractor shall clearly indicate such need. The contractor is not required under this contract to perform any laboratory work.

BACKGROUND

Denison Mines, in co-operation with the Department of Energy, Mines and Resources, is conducting bioleaching of uranium underground. The volumes of leachate encountered in bioleaching are, understandably, enormous and low in concentration. The costs, of pumping millions of gallons per day of such leachate, are very high and the Department wishes to determine the most cost-effective method of recovery of the uranium.

OBJECTIVES

The objectives of the proposed study are:

(a) to draw up suitable conceptual flow sheets for the concentration of bioleachate lixiviant from uranium mines, based on the use of biomass, ion-exchange and reverse osmosis;

(b) to provide technical and economic comparison of the three (3) methods;

(c) to outline the experimental work, if any, that may have to be performed to confirm any assumptions that may have been made; and

(d) to provide preliminary cost estimates based on the above, identify other economic criteria and make appropriate recommendations.

WORK STATEMENT

The contractor shall proceed with the work in stages as detailed below. However, should the studies at any stage, in the opinion of the Scientific Authority, be deemed to be of no practical value, then the investigation shall cease at that point, and no further study of the succeeding stages shall be done.

STAGE I

The contractor shall submit to the Scientific Authority detailed flow sheets for the three (3) processes, namely the use of biomass, ion-exchange and reverse osmosis, supporting wherever necessary the choice and indicating the experimental work that may have to be done to confirm any assumptions that may have been made.

.../2

STAGE II

On approval of the flow sheets by the Scientific Authority, the contractor shall carry out technical and economic evaluation of the three (3) processes. The economic evaluation should include sizing and costing of typical equipment, capital and operating costs and other economic criteria, which in the opinion of the contract are relevant.

In each of the reports issued for the two (2) stages, an essential factor is that the contractor identifies and elaborates upon the selection criteria.

DELIVERABLES

The contractor will issue the phase reports as follows:

Report #1: Flowsheets

Report #2: Technical and Economic Evaluation

Report #3: Final Report

Report #4: Seminar

The contractor shall give a seminar on the work performed on a date and time to be specified by the Scientific Authority.

Draft of all reports should be approved by the Scientific Authority before being submitted in the final form.

Ten (10) copies of the final report, complete with an executive summary, in both English and French, will be delivered to the Scientific Authority on or before March 31, 1986.

Draft report entitled "Evaluation of Biomass for Uranium Recovery from Process Streams Laboratory Results" is attached for reference.

APPENDIX B

CAPITAL COST ESTIMATE DETAILS

CAPITAL COST ESTIMATE DETAILS

SUMMARY

KILBORN (SASKATCHEWAN) LTD.

AREA/CLASS MATRIX

RUN : 07 MAR 86

PROJECT: S283-15
MELIS URANIUM BIOLEACH STUDY

RUN NO. F01

TIME: 17:36

AREA	DESCRIPTION	SITEWORK & IMPROVEMENTS (.1---)	BUILDINGS (.2---)	EQUIPMENT GENERAL (.3---)	MECH EQUIP PROCESS (.4---)	PROCESS PIPING (.5---)	ELECTRICAL POWER (.6---)	INSTRU-MENTATION (.7---)	MINING (.8---)	AREA TOTAL
1000	ION EXCHANGE OPTION		90,292		513,448	138,600	199,175	41,100	108,264	1,090,879
2000	BIOMASS ADSORPTION OPTION				1,531,627	337,000	464,045	122,500	334,720	2,789,891
3000	REVERSE OSMOSIS OPTION				1,855,887	222,700	404,050	111,400	135,200	2,729,238

CAPITAL COST ESTIMATE DETAILS

ION EXCHANGE OPTION

KILBORN (SASKATCHEWAN) LTD.

ESTIMATE DETAILS

PROJECT: S283-15
MELIS URANIUM BIOLEACH STUDY

RUN NO. F01

RUN : 07 MAR 86
TIME: 17:09

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 1000 - ION EXCHANGE OPTION													
2000-01	SURFACE BUILDING C/W BUILDING SERVICES	853	CM	1.43	1,220	35	42,693			50.00	42,650	42,650	85,343
2000-02	SURFACE CONCRETE PAD	11	CM	6.57	72	35	2,529			220.00	2,420	2,420	4,949
2000					1,292		45,222				45,070	45,070	90,292*
4310-01	MIXER- IN - LINE	1	LS						500			500	500
4610-01	AIR COMPRESSOR-100CM/H C/W 15 HP MOTOR	1	EA					11,600.00	11,600			11,600	11,600
4630-01	PUMP-ION EXCHANGE FEED 100 X 75	2	EA					11,200.00	22,400			22,400	22,400
4630-02	MOTORS-15 HP-ION EXCHANGE FEED PUMPS	2	EA					650.00	1,300			1,300	1,300
4630-05	PUMPS-BARREN RECYCLE 100 X 150	2	EA					24,200.00	48,400			48,400	48,400
4630-06	MOTORS-75HP,3600 RPM BARREN RECYCLE PUMPS	2	EA					4,200.00	8,400			8,400	8,400
4630-10	PUMP -BACKWASH 100X75	1	EA					11,200.00	11,200			11,200	11,200
4630-11	MOTOR-10HP BACKWASH PUMP	1	EA					400.00	400			400	400
4630									92,100			92,100	92,100*
4680-01	PUMP-RESIN-U/G 100 X 75	1	EA					13,350.00	13,350			13,350	13,350

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 1000 - ION EXCHANGE OPTION													
4680-02	MOTOR-RESIN PUMP 3 HP	1	EA					250.00	250			250	250
4680-05	PUMP-30% ACID 38 X 25	1	EA					6,150.00	6,150			6,150	6,150
4680-06	MOTOR-2 HP, 30% ACID PUMP	1	EA					200.00	200			200	200
4680-10	PUMP- SPENT ACID 75 X 38	1	EA					13,350.00	13,350			13,350	13,350
4680-11	MOTOR-SPENT ACID 7.5 HP	1	EA					350.00	350			350	350
4680-15	PUMP-RESIN-MILL-PORTABLE TRANSFER 100 X 75	1	EA					13,350.00	13,350			13,350	13,350
4680-16	MOTOR-RESIN PUMP 2HP	1	EA					250.00	250			250	250
4680									47,250			47,250	47,250*
46--									150,950			150,950	150,950*
4820-01	TANK-BIOLEACH SOLUTION HEAD (2500MM DIA.X 2500)	1245	KG					3.25	4,046			4,046	4,046
4820-02	RUBBER LINING-BIOLEACH SOL'N HEAD TANK	25	SM	4.30	108	35	3,763			145.00	3,625	3,625	7,388
4820-05	ION EXCHANGE COLUMNS 2000MM DIA X 3250 (6 REQ'D)	12912	KG					3.25	41,964			41,964	41,964
4820-06	RUBBER LINING-ION EXCHANGE COLUMNS	174	SM					188.00	32,712			32,712	32,712

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 1000 - ION EXCHANGE OPTION													
4820-07	INTERNALS-ION EXCHANGE COLUMNS	1	LS						15,000			15,000	15,000
4820-10	TANK-BARREN RECYCLE 2500MM DIA. X 2500	1245	KG					3.25	4,046			4,046	4,046
4820-11	RUBBER LINING-BARREN RECYCLE TANK	25	SM	4.30	108	35	3,763			145.00	3,625	3,625	7,388
4820-15	RESIN TRAP 2000MM DIA.X 4000	1494	KG					3.25	4,856			4,856	4,856
4820-16	RUBBER LINING RESIN TRAP	30	SM					188.00	5,640			5,640	5,640
4820-20	TANK-RESIN HOLDING 2600MM DIA X 2600	1345	KG					3.25	4,371			4,371	4,371
4820-21	RUBBER LINING-RESIN HOLDING TANK	27	SM	4.30	116	35	4,064			145.00	3,915	3,915	7,979
4820-25	ELUTION COLUMN 2500MM DIA X 3750	3042	KG					3.25	9,887			9,887	9,887
4820-26	RUBBER LINING-ELUTION COLUMN	41	SM	4.30	176	35	6,171			145.00	5,945	5,945	12,116
4820-27	INTERNALS-ELUTION COLUMN	1	LS						3,000			3,000	3,000
4820-30	REGENERATION COLUMN 2500MM DIA X 3750	3042	KG					3.25	9,887			9,887	9,887
4820-31	RUBBER LINING-REGEN. COLUMN	41	SM	4.30	176	35	6,171			145.00	5,945	5,945	12,116

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 1000 - ION EXCHANGE OPTION													
4820-32	INTERNALS - REGENERATION COLUMN	1	LS						3,000			3,000	3,000
4820-35	TANK-30% H2SO4 4500MM DIA X 5500	5478	KG					3.25	17,804			17,804	17,804
4820-36	RUBBER LINING -H2SO4 TANK	110	SM	4.30	473	35	16,555			145.00	15,950	15,950	32,505
4820					1,157		40,485		156,212		39,005	195,217	235,701*
4910-01	TANKER-RESIN -MOBILE 2000MM DIA X 4500	1574	KG					3.25	5,181			5,181	5,181
4910-02	RUBBER LINING-RESIN TANKER	32	SM					188.00	6,016			6,016	6,016
4910-03	WHEELED TRAILER	1	LS						2,500			2,500	2,500
4910-04	FIELD MATERIAL - 10% OF H.O.MATERIAL	1	LS								32,100	32,100	32,100
4910-05	MECHANICAL INSTALLATION 25% OF H.O.MATERIAL	1	LS		2,300	35	80,500						80,500
4910					2,300		80,500		13,697		32,100	45,797	126,297*
4---					3,457		120,985		321,358		71,105	392,463	513,448*
5000-01	PROCESS PIPING - 27% OF INSTALLED MECHANICAL	1	LS		2,377	35	83,200				55,400	55,400	138,600

KILBORN (SASKATCHEWAN) LTD.

ESTIMATE DETAILS

PROJECT: S283-15
MELIS URANIUM BIOLEACH STUDY

RUN NO. F01

RUN : 07 MAR 86
TIME: 17:09

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 1000 - ION EXCHANGE OPTION													
6000-01	UNDERGROUND POWER SUPPLY	1	LS		546	35	19,110				29,890	29,890	49,000
6000-02	LIGHTING	1	LS		165	35	5,775				5,800	5,800	11,575
6000-03	EQUIPMENT ELECTRICAL 27% OF INSTALLED MECHANICAL	1	LS		1,980	35	69,300				69,300	69,300	138,600
6000					2,691		94,185				104,990	104,990	199,175*
7000-01	INSTRUMENTATION - 8% OF INSTALLED MECHANICAL	1	LS		294	35	10,300				30,800	30,800	41,100
8000-01	UNDERGROUND EXCAVATION	1098	M	1.86	2,039	35	71,365			25.00	27,450	27,450	98,815
8000-02	UNDERGROUND CONCRETE PAD	13	CM	8.71	157	35	5,490			220.00	3,960	3,960	9,450
8000					2,196		76,854				31,410	31,410	108,264*
AREA 1000 TOTAL					12,307		430,746		321,358		338,775	660,133	1,090,879*

CAPITAL COST ESTIMATE DETAILS

BIOMASS ADSORPTION OPTON

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 2000 - BIOMASS ADSORPTION OPTION													
4150-01	CENTRIFUGE-LOADED 610 X 2490	1	EA					286,000	286,000			286,000	286,000
4150-02	MOTOR-100 HP CENTRIFUGE	1	EA					8,000.00	8,000			8,000	8,000
4150-05	CENTRIFUGE-DESORBED 432 X 1270	1	EA					137,000	137,000			137,000	137,000
4150-06	MOTOR-50 HP DESORBED CENTRIFUGE	1	EA					6,000.00	6,000			6,000	6,000
4150									437,000			437,000	437,000*
4240-01	DUST COLLECTOR-LIME SILO	1	EA					4,700.00	4,700			4,700	4,700
4240-05	DUST COLLECTOR-BICARB SILO	1	EA					4,700.00	4,700			4,700	4,700
4240									9,400			9,400	9,400*
4300-01	BIN VIBRATOR-LIME SILO 1200 MM DIA	1	EA					9,450.00	9,450			9,450	9,450
4300-05	BIN VIBRATOR -BICARB. SILO 1200MM DIA	1	EA					9,450.00	9,450			9,450	9,450
4300									18,900			18,900	18,900*
4310-01	AGITATOR- PH ADJUSTMENT TANKS	2	EA					7,700.00	15,400			15,400	15,400
4310-02	MOTORS-7.5 HP AGITATORS PH ADJUSTMENT	2	EA					350.00	700			700	700

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 2000 - BIOMASS ADSORPTION OPTION													
4310-05	AGITATOR-SLURRY HOLDING TANK	1	EA					3,000.00	3,000			3,000	3,000
4310-06	MOTOR-AGITATOR SLURRY HOLDING TANK	1	EA					200.00	200			200	200
4310-10	AGITATOR - BIOADSORPTION TANK	1	EA					7,700.00	7,700			7,700	7,700
4310-11	MOTOR-7.5 HP AGITATOR BIOADSORPTION TANK	1	EA					350.00	350			350	350
4310-15	AGITATOR-ELUTION TANK	1	EA					3,500.00	3,500			3,500	3,500
4310-16	MOTOR-AGITATOR ELUTION TANK 5 HP	1	EA					300.00	300			300	300
4310-20	AGITATOR-BICARB. TANK	1	EA					4,700.00	4,700			4,700	4,700
4310-21	MOTOR-5HP AGITATOR BICARB.TANK	1	EA					300.00	300			300	300
4310									36,150			36,150	36,150*
4370-01	FLOCCULANT EDUCTOR	1	LS						100			100	100
43--									55,150			55,150	55,150*
4430-01	SCREW FEEDER-LIME C/W MOTOR	1	LS						4,000			4,000	4,000
4430-05	SCREW FEEDER-BICARB. C/W MOTOR	1	LS						4,000			4,000	4,000

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 2000 - BIOMASS ADSORPTION OPTION													
4430-10	SCREW FEEDER- BIOMASS C/W MOTOR	1	LS						4,000			4,000	4,000
4430									12,000			12,000	12,000*
4610-01	AIR COMPRESSOR C/W RECEIVER & 15 HP MOTOR	1	EA					11,600.00	11,600			11,600	11,600
4630-01	PUMPS-BIOADSORPTION FEED 100 X 75	2	EA					11,200.00	22,400			22,400	22,400
4630-02	MOTORS-BIOADSORPTION FEED PUMPS 10 HP	2	EA					400.00	800			800	800
4630-05	PUMP-CLARIFIER U/F 38 X 25	1	EA					5,900.00	5,900			5,900	5,900
4630-06	MOTOR-CLARIFIER U/F PUMP 1HP	1	EA					200.00	200			200	200
4630-10	PUMP-PRECIPITATE SLUDGE 19 X 19	1	EA					3,900.00	3,900			3,900	3,900
4630-11	MOTOR-5 HP PRECIPITATE SLUDGE PUMP	1	EA					300.00	300			300	300
4630-15	PUMP-BIOMASS U/F 75 X 50	1	EA					9,700.00	9,700			9,700	9,700
4630-16	MOTOR-BIOMASS U/F PUMP 5 HP	1	EA					300.00	300			300	300

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 2000 - BIOMASS ADSORPTION OPTION													
4630-20	PUMPS-BARREN RECYCLE 100 X 50	2	EA					24,200.00	48,400			48,400	48,400
4630-21	MOTOR-75 HP 3600 RPM BARREN RECYCLE PUMPS	2	EA					4,200.00	8,400			8,400	8,400
4630-25	PUMP-ELUTION 75 X 38	1	EA					9,450.00	9,450			9,450	9,450
4630-26	MOTOR-ELUTION PUMP 5 HP	1	EA					300.00	300			300	300
4630									110,050			110,050	110,050*
4680-01	PUMP-BICARB.FEED 38 X 25	1	EA					5,300.00	5,300			5,300	5,300
4680-02	MOTOR-3 HP-BICARB. FEED PUMP	1	EA					250.00	250			250	250
4680-05	PUMP-FLOCCULANT 38 X 32	1	EA					3,660.00	3,660			3,660	3,660
4680-06	MOTOR-FLOCC. PUMP 0.75HP	1	EA					200.00	200			200	200
4680									9,410			9,410	9,410*
46--									131,060			131,060	131,060*
4820-01	TANK-BIOADSORPTION FEED 2500MM DIA X 2500	1245	KG					3.25	4,046			4,046	4,046
4820-02	RUBBER LINING-BIOADSORPTION FEED TANK	25	SM	4.30	108	35	3,763			145.00	3,625	3,625	7,388

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 2000 - BIOMASS ADSORPTION OPTION													
4820-05	TANKS- PH ADJUSTMENT 3250MM DIA X 3250 (2 REQD)	4184	KG					3.25	13,598			13,598	13,598
4820-06	RUBBER LINING- PH ADJUSTMENT TANKS(2 REQD)	84	SM	4.30	361	35	12,642			145.00	12,180	12,180	24,822
4820-10	TANK-PRECIPITATE CLARIFIER-6500MM DIA	5478	KG					3.25	17,804			17,804	17,804
4820-11	RUBBER LINING- PRECIPITATE CLARIFIER TANK	110	SM	4.30	473	35	16,555			145.00	15,950	15,950	32,505
4820-12	S/S MECHANISM C/W BRIDGE & MOTOR	1	EA					95,600.00	95,600			95,600	95,600
4820-15	TANK-SLURRY HOLDING 1200MM DIA X 1200	249	KG					3.25	809			809	809
4820-16	RUBBER LINING-SLURRY HOLDING TANK	5	SM					188.00	940			940	940
4820-20	TANK-BIOADSORPTION 4000MM DIA X 4000	3137	KG					3.25	10,195			10,195	10,195
4820-21	RUBBER LINING- BIOADSORPTION TANK	63	SM	4.30	271	35	9,482			145.00	9,135	9,135	18,617
4820-25	TANK-BIOMASS CLARIFIER 10000MM DIA	9810	KG					3.25	31,883			31,883	31,883

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 2000 - BIOMASS ADSORPTION OPTION													
4820-26	RUBBER LINING-BIOMASS CLARIFIER TANK	197	SM	4.30	847	35	29,649			145.00	28,565	28,565	58,214
4820-27	S/S MECHANISM C/W BRIDGE & MOTOR	1	EA					112,100	112,100			112,100	112,100
4820-30	TANK-BARREN RECYCLE 2500MM DIA X 2500	1245	KG					3.25	4,046			4,046	4,046
4820-31	RUBBER LINING-BARREN RECYCLE TANK	25	SM	4.30	108	35	3,763			145.00	3,625	3,625	7,388
4820-35	TANK-ELUTION 1600MM DIA X 1600	349	KG					3.25	1,134			1,134	1,134
4820-36	RUBBER LINING-ELUTION TANK	7	SM					188.00	1,316			1,316	1,316
4820-40	TANK-ELUATE SUMP 1200MM DIA X 1200	249	KG					3.25	809			809	809
4820-41	RUBBER LINING-ELUATE SUMP TANK	5	SM					188.00	940			940	940
4820-45	LIME SILO-2000MM DIA X 4000	2056	KG					3.25	6,682			6,682	6,682
4820-50	BICARBONATE SILO 2000MM DIA X 4000	2056	KG					3.25	6,682			6,682	6,682
4820-55	TANK-BICARBONATE 2500MM X 2500	1245	KG					3.25	4,046			4,046	4,046

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 2000 - BIOMASS ADSORPTION OPTION													
4820-60	TANK-FLOCCULANT 2000MM DIA X 2000	797	KG					3.25	2,590			2,590	2,590
4820-61	RUBBER LINING-FLOCC. TANK	16	SM					188.00	3,008			3,008	3,008
4820-65	BIN-BIOMASS STORAGE 2500MM DIA X 4000	5471	KG					3.25	17,781			17,781	17,781
4820					2,167		75,852		336,010		73,080	409,090	484,942*
4910-01	FIELD MATERIAL-10% OF H.O. MATERIAL	1	LS								98,100	98,100	98,100
4920-01	MECHANICAL INSTALLATION 31 % H.O. MATERIAL	1	LS		8,685	35	303,975						303,975
49--					8,685		303,975				98,100	98,100	402,075*
4---					10,852		379,827		980,620		171,180	1,151,800	1,531,627*
5000-01	PROCESS PIPING-22 % OF INSTALLED MECHANICAL	1	LS		5,777	35	202,200				134,800	134,800	337,000
6000-01	UNDERGROUND POWER SUPPLY	1	LS		637	35	22,295				49,700	49,700	71,995
6000-02	U/G LIGHTING	1	LS		130	35	4,550				4,600	4,600	9,150
6000-03	EQUIPMENT ELECTRICAL 25 % OF INSTALLED MECHANICAL	1	LS		5,471	35	191,500				191,400	191,400	382,900
6000					6,238		218,345				245,700	245,700	464,045*

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 2000 - BIOMASS ADSORPTION OPTION													
7000-01	INSTRUMENTATION-B % OF INSTALLED MECHANICAL	1	LS		874	35	30,600				91,900	91,900	122,500
8000-01	UNDERGROUND EXCAVATION	3581	CM	1.86	6,661	35	233,123			25.00	89,525	89,525	322,648
8000-02	UNDERGROUND CONCRETE PAD	23	CM	8.71	200	35	7,012			220.00	5,060	5,060	12,072
8000					6,861		240,135				94,585	94,585	334,720+
AREA 2000 TOTAL					30,603		1,071,107		980,620		736,165	1,716,785	2,789,891+

CAPITAL COST ESTIMATE DETAILS

REVERSE OSMOSIS OPTION

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 3000 - REVERSE OSMOSIS OPTION													
4130-01	BAG FILTER-2 SM CAPACITY	1	EA					9,800.00	9,800			9,800	9,800
4310-01	AGITATOR REAGENT MAKE-UP TANK	1	EA					6,000.00	6,000			6,000	6,000
4310-02	MOTOR-5 HP-REAGENT MAKE-UP AGITATOR	1	EA					300.00	300			300	300
4310									6,300			6,300	6,300*
4380-01	REVERSE OSMOSIS UNIT (62 MODULES)	1	EA					1,000,000	1,000,000			1,000,000	1,000,000
43--									1,006,300			1,006,300	1,006,300*
4610-01	AIR COMPRESSOR-100CM/HR C/W RECEIVER & 15 HP MOTOR	1	EA					11,600.00	11,600			11,600	11,600
4630-01	PUMP-SAND FILTER FEED 100 X 75	2	EA					11,200.00	22,400			22,400	22,400
4630-02	MOTOR-20 HP SAND FILTER FEED PUMP	2	EA					800.00	1,600			1,600	1,600
4630-05	PUMP-BAG FILTER FEED 100 X 75	2	EA					11,200.00	22,400			22,400	22,400
4630-06	MOTOR-BAG FILTER FEED PUMP 20 HP	2	EA					800.00	1,600			1,600	1,600
4630-10	PUMPS-R.O. FEED 75 X 150	2	EA					82,100.00	164,200			164,200	164,200

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	H.O. MATERIAL -- TOTAL	- FIELD MATERIAL - UNT CST	FIELD MATERIAL - TOTAL	TOTAL MAT'L	TOTAL COST
AREA 3000 - REVERSE OSMOSIS OPTION													
4630-11	MOTORS-R.O.FEED PUMPS 250 HP 3600RPM	2	EA					15,500.00	31,000			31,000	31,000
4630-15	PUMPS-PERMEATE RECYCLE 75 X 100	2	EA					24,200.00	48,400			48,400	48,400
4630-16	MOTORS-PERMEATE RECYCLE PUMP 60 HP	2	EA					3,500.00	7,000			7,000	7,000
4630-20	PUMP-BACKWASH 100 X 75	1	EA					11,200.00	11,200			11,200	11,200
4630-21	MOTOR -BACKWASH PUMP 15 HP	1	EA					650.00	650			650	650
4630									310,450			310,450	310,450*
4680-01	PUMP-REAGENT FEED 38 X 50	1	EA					7,700.00	7,700			7,700	7,700
4680-02	MOTOR-REAGENT FEED PUMP 20 HP 3600RPM	1	EA					1,000.00	1,000			1,000	1,000
4680									8,700			8,700	8,700*
46--									330,750			330,750	330,750*
4820-01	TANK-SAND FILTER FEED 2500MM DIA X 2500	1245	KG					3.25	4,046			4,046	4,046
4820-02	RUBBER LINING-SAND FILTER FEED TANK	25	SM	4.30	108	35	3,763			145.00	3,625	3,625	7,388

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 3000 - REVERSE OSMOSIS OPTION													
4820-05	TANK-BAG FILTER FEED 2500MM DIA X 2500	1245	KG					3.25	4,046			4,046	4,046
4820-06	RUBBER LINING-FILTER FEED TANK	25	SM	4.30	108	35	3,763			145.00	3,625	3,625	7,388
4820-10	TANK-R.O. FEED 2500MM DIA X 2500	1245	KG					3.25	4,046			4,046	4,046
4820-11	RUBBER LINING-R.O. FEED TANK	25	SM	4.30	108	35	3,763			145.00	3,625	3,625	7,388
4820-15	TANK-PERMEATE RECYCLE 3000MM DIA X 3000	1793	KG					3.25	5,827			5,827	5,827
4820-16	RUBBER LINING-PERMEATE RECYCLE TANK	36	SM	4.30	155	35	5,418			145.00	5,220	5,220	10,638
4820-20	TANK-REAGENT MAKE-UP 3000MM DIA X 3000	1793	KG					3.25	5,827			5,827	5,827
4820-21	RUBBER LINING-REAGENT MAKE-UP TANK	36	SM	4.30	155	35	5,418			145.00	5,220	5,220	10,638
4820-25	TANK-CONCENTRATE SOLUTION 1500MM DIA X 1500	448	KG					3.25	1,456			1,456	1,456
4820-26	RUBBER LINING-CONCENTRATE SOLUTION TANK	9	SM					188.00	1,692			1,692	1,692
4820					632		22,124		26,941		21,315	48,256	70,380*

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 3000 - REVERSE OSMOSIS OPTION													
4830-01	SAND FILTERS 2000MM DIA X 3000 (3 REQD)	6009	KG					3.25	19,529			19,529	19,529
4830-02	RUBBER LINING-SAND FILTERS (3REQD)	81	SM					188.00	15,228			15,228	15,228
4830-03	INTERNALS-3 SAND FILTERS	1	LS						7,500			7,500	7,500
4830									42,257			42,257	42,257*
48--					632		22,124		69,199		21,315	90,514	112,637*
4910-01	FIELD MATERIAL-10% OF H.O.MATERIAL	1	LS								141,600	141,600	141,600
4920-01	MECHANICAL INSTALLATION 18% OF H.O. MATERIAL	1	LS		7,280	35	254,800						254,800
49--					7,280		254,800				141,600	141,600	396,400*
4---					7,912		276,924		1,416,049		162,915	1,578,964	1,855,887*
5000-01	PROCESS PIPE-12% OF INSTALLED MECHANICAL	1	LS		3,817	35	133,600				89,100	89,100	222,700
6000-01	UNDERGROUND POWER SUPPLY	1	LS		770	35	26,950				71,050	71,050	98,000
6000-02	LIGHTING	1	LS		130	35	4,550				4,600	4,600	9,150
6000-03	EQUIPMENT ELECTRICAL 16% OF INSTALLED MECHANICAL	1	LS		4,243	35	148,500				148,400	148,400	296,900
6000					5,143		180,000				224,050	224,050	404,050*

CLSS ITM	DESCRIPTION	QUAN	UN	UNIT MHRS	TOTAL MNHRS	MH RT	TOTAL LABOR	-- H.O. MATERIAL -- UNT CST	TOTAL	- FIELD MATERIAL - UNT CST	TOTAL	TOTAL MAT'L	TOTAL COST
AREA 3000 - REVERSE OSMOSIS OPTION													
7000-01	INSTRUMENTATION-6% OF INSTALLED MECHANICAL	1	LS		797	35	27,900				83,500	83,500	111,400
8000-01	UNDERGROUND EXCAVATION	1372	CM	1.85	2,538	35	88,837			25.00	34,300	34,300	123,137
8000-02	UNDERGROUND CONCRETE PAD	23	CM	8.70	200	35	7,003			220.00	5,060	5,060	12,063
8000					2,738		95,840				39,360	39,360	135,200*
AREA 3000 TOTAL					20,408		714,284		1,416,049		598,925	2,014,974	2,729,257*